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Assessment of Three Technologies for the Treatment of Cooling Tower Blowdown

Bechtel National, Inc, San Francisco, CA

Prepared for

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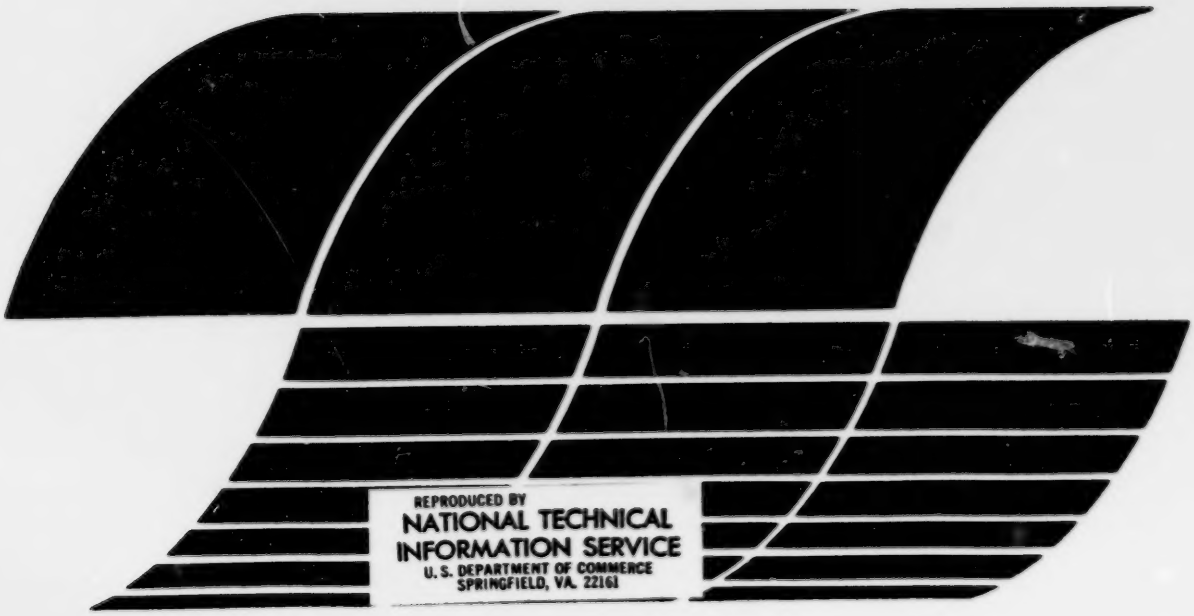
Interagency Energy/Environment R&D Program Report



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16. ABSTRACT

The report gives results of analyses of three methods for treating cooling tower blowdown: vapor compression evaporation (VCE), reverse osmosis (RO), and vertical tube foaming evaporation (VTFE). The two evaporative processes produce pure water (approximately 10 ppm dissoved solids). RO produces water of lower purity (about 500 ppm) but adequate for many uses in the power plant or for return to the cooling tower. VCE has been used successfully in commercial power plants; the evaporative processes have no plants in operation on cooling tower blowdown. Plant investment is strongly influenced by the cost of lined ponds required to evaporate the treatment plant blowdown and store the residual salts. Consequently, the RO plant investment is almost in the range of that of VTFE and VCE plants. In total capital plus operating costs, VTFE has a distinct advantage over RO and VCE becuase its source of energy is waste heat to which a zero value has been assigned. The VTFE economics is based on vertical tube evaporator experience (without adding surfactant). If field pilot tests substantiate previous laboratory results, the economics of the VTFE may prove to be even more favorable.

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Industrial Processes	Electric Power Plants	Stationary Sources	13H	10B
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by

E. H. Houle, A. N. Rogers, M. C. Weekes,
S. C. May and V. C. Van der Mast

Bechtel National, Inc.
P. O. Box 3965
San Francisco, CA 94119

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EPA Project Officer: Michael C. Osborne

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, NC 27711

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FOREWORD

The objective of this study is to examine methods for reducing the volume of blowdown from cooling towers in order to decrease the cost of ponding or disposal of the blowdown while at the same time recovering water of a quality capable of reuse at some point within the plant.

This project was conducted under the sponsorship of the Industrial Environmental Research Laboratory of the U.S. Environmental Protection Agency under Contract No. 68-02-2616, entitled "Monitoring of the Vertical Tube Foam Evaporation Demonstration (VTFE-D) and the Assessment of Various Technologies for the Treatment of Cooling Tower Blowdown." The assigned work scope embraces two tasks:

1) Monitor progress, assist in formulating test plan, prepare rough cost evaluation of VTFE-D

2) Compare economic and energy efficiency merits of VTFE, reverse osmosis, and vapor compression evaporation.

ACKNOWLEDGMENTS

Several equipment manufacturers kindly contributed case histories from the commercial use of their plants and equipment and operating costs.

Information on the performance and costs of commercial vertical tube evaporator plants was supplied by Mr. R. H. Hedrick of the Goslin-Birmingham Division of Envirotech, Mr. Don Nelson of Struthers Scientific and International Corporation, and Mr. Malcolm Coston of the Swenson Division of the Whiting Corporation.

Technical and economic data on vapor-compression evaporators were furnished by Mr. Wayne E. Springer of Resources Conservation Company.

Cost and performance data on reverse osmosis systems were supplied by Mr. D. C. Brandt of DuPont's Permasep Division, Ms. Mary Jenkins of Dow Chemical Company's Membrane Division, Ms. Amy Knapp of Polymetrics, and Mr. I. Nusbaum of UOP's Fluid Systems.

Dr. H. H. Sephton, Principal Investigator of the University of California Sea Water Conversion Laboratory and the inventor of the VTFE concept, cooperated in the preparation of the test plan for the VTFE shakedown tests and provided details and data for this report.

SECTION 1

INTRODUCTION

This study, performed under Contract No. 68-02-2616 of the Emissions/Effluent Technology Branch of the U.S. Environmental Protection Agency, was directed toward the concentration of cooling tower blowdown of power plants.

Lack of adequate supplies of cooling water and increasing concern over thermal pollution of streams and lakes have forced the management of many power stations, especially in western states, to consider replacing once-through cooling systems with cooling towers. To conserve water, dry cooling towers can be used, but at the expense of increasing turbine back pressure, which lowers power plant efficiency. Consequently, either the wet or the wet-dry combination cooling tower is preferred. For either of these choices, evaporating a fraction of the water in the coolant cycle produces a saline blowdown stream. This stream, in the context of a zero discharge policy being more seriously considered at the present time, may not be discharged into surface water supplies. Ponding of these aqueous wastes is costly, because of the necessary land area and because of the expensive liners required to prevent seepage into the soil during prolonged storage.

It is apparent that the volume of material to be ponded (and therefore the cost of ponding) can be reduced by concentrating the aqueous waste stream discharged from cooling towers. Water can be recovered from the cooling tower blowdown waste stream to yield a product that could be used in several locations in a power plant. This could make the plant less dependent on fresh water supplies, an important consideration in some water-short regions of the United States. Possible uses are for the scrubbers of fossil fueled plants, ash sluicing in coal burning plants, pump gland seal water, boiler makeup, and recycle to the cooling tower loop (one of the options considered in this study).

Several methods are available for treating cooling tower blowdown streams, including:

- Reverse osmosis
- Vapor compression evaporation
- Vertical tube foaming evaporation
- Conventional softening
- Multistage flash distillation
- Thermal softening

Because of budgetary restrictions, it was agreed that this study would be limited to three of the above techniques. The selection of the candidate processes was based on the fact that the desalting-related technologies produce a relatively pure product which could generally be used as a recyle feed at a number of points in the plant; the softening techniques produce a product with more restricted uses, due to the residual dissolved salts. This fact, and the fact that multistage flash distillation technology is well known, led to an agreement to direct the present study toward the first three of these processes.

The success of these processes or any other technique for waste water concentration depends on preventing excessive scale formation through controlling the chemistry of the feed stream. The significance of scaling and the need to avoid scale formation is discussed under the respective processes. For a brief discussion of basic water chemistry, please refer to Appendix A.

In the following sections, each of these processes is examined from the standpoint of technology, operating experience, and cost. The processes are then compared, with a discussion of their relative merits and shortcomings. Finally, there is a discussion on the advantages of a combined process in which the reject stream from a reverse osmosis plant is further concentrated by means of an evaporator-crystallizer.

Bechtel cooperated in the planning of shakedown tests on the pilot plant vertical tube foaming evaporator to be performed at the Sea Water Conversion Laboratory of the University of California. Details of the test plan are included in this report.

SECTION 2

CONCLUSIONS

Table 1 compares the three candidate processes for the concentration of cooling tower blowdown. It appears at present that no process is definitely superior to the other two in all respects. Future pilot plant work and industrial experience may alter this conclusion. For instance, the current field testing of the VTFE would, if successful, establish the reliability of this process, for which there is no basis at present. Costs would also be lower if substantially higher heat transfer rates could be demonstrated than conservatively estimated in this study.

Although not specifically studied, a combination process may be superior to any one of the individual processes examined in this report. An interesting possibility is the pre-concentration of cooling tower blowdown by a membrane process such as RO or ED, sending the reject stream for final volume reduction to a distillation plant such as VTFE, vapor compression, or multistage flash.

TABLE 1. COMPARISON OF PROCESSES[1]

	Reverse Osmosis	Vertical Tube Foaming Evaporation	Vapor Compression Evaporation
Economics			
Plant investment 1st quarter '79 (million dollars)	13.9	15.9	15.2
Annual cost (million dollars per year)[2]			
Capital cost	1.7	2.0	1.9
O&M cost	1.2	0.6	2.9
Total annual cost	2.9	2.6	4.8
Cost/kgal of feed	3.28	3.41	6.06
Cost/kgal of purified water	4.37	3.48	6.18
Energy consumption[3]	Lowest	Low (if steam is "free")	High
Ponding requirements[4]	Large	Moderate	Moderate
Long-term related operating experience on cooling tower blowdown	Minimal	None	Considerable and successful
Modular construction of sub-units of each train	Standard procedure	Not practical	Not practical
Sensitivity to upsets in pretreatment	Very sensitive	Minor	Minor
Product quality	Fair (475 ppm)	High (10 ppm)	High (10 ppm)

(1) See Table 9 for details.
(2) Basis: 30-year plant life, ponds at $50,000/acre, electricity at 4¢/kWh, 12.5% fixed charge rate. See body of this report for a discussion of effects of variations in these assumptions.
(3) Energy cost is included in O&M cost above.
(4) Ponding cost is included in the plant investment and the calculation of annual cost above.

SECTION 3

RECOMMENDATIONS

The shakedown tests of the vertical tube foam evaporator (VTFE) at the Sea Water Conversion Laboratory (SWCL) are being initiated as this report is being completed. Consequently, no conclusion can be reached at this time on the reliability and economy of the process under actual field conditions. It is recommended, therefore, that both shakedown runs at SWCL and field tests at a candidate power plant be monitored and that the resulting data be used to update the calculations and conclusions of this report.

Simultaneously, the "first cut" analyses presented here should be refined to include a more detailed consideration of the overall water balance of a typical power plant. Credit should be taken for the value of product water of different degrees of purity produced by each of the three wastewater concentration processes. Water not required for high quality usage (such as boiler feed water) can be returned to the cooling tower loop, thereby decreasing the size and cost of the wastewater concentration process, as demonstrated in this report.

In future work, use of electrodialysis (ED) should be considered; under some circumstances, it is competitive with reverse osmosis (RO) with respect to both economics and to reducing the volume of cooling tower blowdown. At the same time, work should proceed on a study of a two-step process. Such a study should involve a membrane plant, RO or ED, of which the reject stream is further concentrated by an evaporative process such as VTFE, vapor compression, or multistage flash evaporation. The combined process will not only deliver a stream of purified water for boiler makeup but will sharply reduce the volume of reject brine requiring ponding with a potential reduction in overall cost.

Another attractive alternative for future consideration is the use of a softening process such as lime or thermal softening. A portion of its product can be further purified for boiler feed water while the balance of the product, which is comparatively low cost, can be used for cooling tower makeup, scrubber water, and other purposes not requiring such high purity.

Section 4

ANALYSIS AND COMPARISON OF PROCESSES

In this section, three processes for concentrating cooling tower blowdown are analyzed: reverse osmosis (RO), vapor compression (VC), and vertical tube foaming evaporation (VTFE). The last two processes are operated to allow the presence of substantial amounts of suspended solids in the brine being concentrated. Reverse osmosis, however, must be operated in a scale-free manner, and therefore, requires pretreatment of the cooling tower blowdown (described in detail in Subsection 4.1.3).

It should be noted that at the time of writing of this report, no VTFE plant was in operation, nor had any data been generated by the pilot plant. The design parameters and economic analysis, therefore, were based on commercial experience with vertical tube evaporators (VTE) in accordance with EPA instructions. All references to the VTFE design in this report refer to studies on the VTE and will require updating when the results of the VTFE pilot plant field tests become available.

Since reverse osmosis and its costs are sensitive to the composition of the feed stream, two water compositions were examined: one limited by silica solubility (Water A) and one by $CaSO_4$ solubility (Water B). In each case, the ion content of a commercial power plant cooling tower blowdown was approximated. For the distillation processes, where costs are not sensitive to moderate changes in composition, only Water A was considered, and softening of the blowdown was eliminated. The compositions of water A and Water B are shown in Tables 2 and 3. The second column in each table lists the compositions after pretreatment.

The conceptual design in each case is based on a typical 700 MWe fossil fueled power plant. The cooling tower is assumed to operate at four cycles of concentration. That is, three-fourths of the water content of the makeup stream is evaporated, resulting in a blowdown in which the total dissolved solids concentration is four times that in the makeup.

TABLE 2. WATER A: CONCENTRATION OF WATER LIMITED BY SILICA SCALING

	Composition Expressed as ppm of Ion	
	Cooling Tower Blowdown	Blowdown After Softening and pH Adjustment
Na	1,320	2,018
Ca	400	50
Mg	400	236
Cl	208	208
SO_4	5,000	5,027
HCO_3	44	43
SiO_2	150	37
Sum of ions	7,522	7,619
pH	7.3	6.0
T	80°F	80°F

TABLE 3. WATER B: CONCENTRATION OF WATER LIMITED BY CALCIUM SULFATE SCALING

	Composition Expressed as ppm of Ion	
	Cooling Tower Blowdown	Blowdown After Softening and pH Adjustment
Na	710	1,265
Ca	533	50
Mg	193	174
Cl	266	266
SO_4	3,820	3,847
HCO_3	35	43
SiO_2	5	5
Sum of ions	5,562	5,650
pH	6.7	6.0
T	80°F	80°F

Each of the processes produces a reject stream requiring ponding or disposal. In this study, it is assumed that the reject brine is discharged to plastic-lined solar evaporation ponds. These ponds are economically attractive only in locations where evaporation exceeds precipitation. In regions of high rainfall, the ponds require covers, which add substantially to their cost.

This report was directed primarily toward the arid regions of the western United States, where brine disposal is a serious problem and the conservation of cooling tower water in a power plant is highly desirable. In these regions, net evaporation (evaporation minus rainfall) is high. The net evaporation from brine in Arizona, for example, is reported to be 5.7 feet per year.(Ref. 1) The annual evaporation rate, however, is not a constant at any particular location but is influenced, among other things, by the height of the berm above the liquid level in the pond. In this study, therefore, the conservative value of 4.0 feet of net evaporation per year was used. The usable depth of the ponds was assumed to be 10 feet.

For the evaporation processes, for which the reject flow was low, a pond area was assumed adequate to hold the crystallized salts deposited from 30 years' flow of reject brine. The surface area was found to be far in excess of that required to evaporate all the contained water at an evaporation rate of 4.0 feet per year. For the RO system, however, additional pond area was required to evaporate the large reject flow and contain the unevaporated water at the end of 30 years.

In the economic section of each process, the cost of solar evaporation ponds was calculated on the basis of $40,000, $50,000, and $60,000 per acre, and the pond costs were added to the installed costs of the respective plants.

There are several alternatives to ponding, among them deep well injection. However, injection requires extensive pretreatment of the aqueous waste stream to ensure freedom from suspended matter and from any species that may tend to crystallize or precipitate at downhole conditions. A careful analysis of all related costs is required before deep well injection is seriously considered.

The selection of a hot desert region in the western United States as the basis for this study led to the assumption of a comparatively high cooling tower temperature and, hence, to a turbine exhaust at 125°F as the heat source to the VTFE plant. In colder climates, the turbine exhaust temperature would be much lower. Since the cooling tower temperature would drop by the same amount, the thermal driving force across the VTFE would be unchanged. However, the VTFE design and equipment costs would then reflect the increase in vapor volume and the decrease in heat transfer coefficient resulting from evaporation at the lower temperature. Alternatively, the VTFE could still be operated at the 125°F steamside temperature assumed in this study by supplying the process with extraction steam from the turbine. In that case, the thermal energy supplied to the VTFE could no longer be considered "free," as assumed here.

4.1 REVERSE OSMOSIS

4.1.1 Design Basis

Precipitation of scale in a reverse osmosis module may completely impede the functioning of the membrane in a very short time. Therefore, it is essential to treat the feed to the RO plant to reduce the concentration of scale formers to the point where their solubilities will not be exceeded in the reject stream, and in particular in the boundary layer between reject brine and membrane, where the concentrations are a maximum. For example, if 75 percent of the water is recovered in purified form from the feed stream, the reject stream then represents only one-fourth of the original volume and thus contains roughly four times the initial concentrations of contaminants.

Consequently, the process design in Figure 1 performs the following:

- Reduces the calcium concentration in the cooling tower blowdown to a value that will avoid $CaSO_4$ scale formation in the reverse osmosis reject stream
- Precipitates Mg as the hydroxide to carry with it sufficient SiO_2 to prevent silica scaling in the reverse osmosis modules
- Adjusts the pH of the softened water to ensure a pH below the Langelier Index for $CaCO_3$ precipitation in the reverse osmosis reject stream.

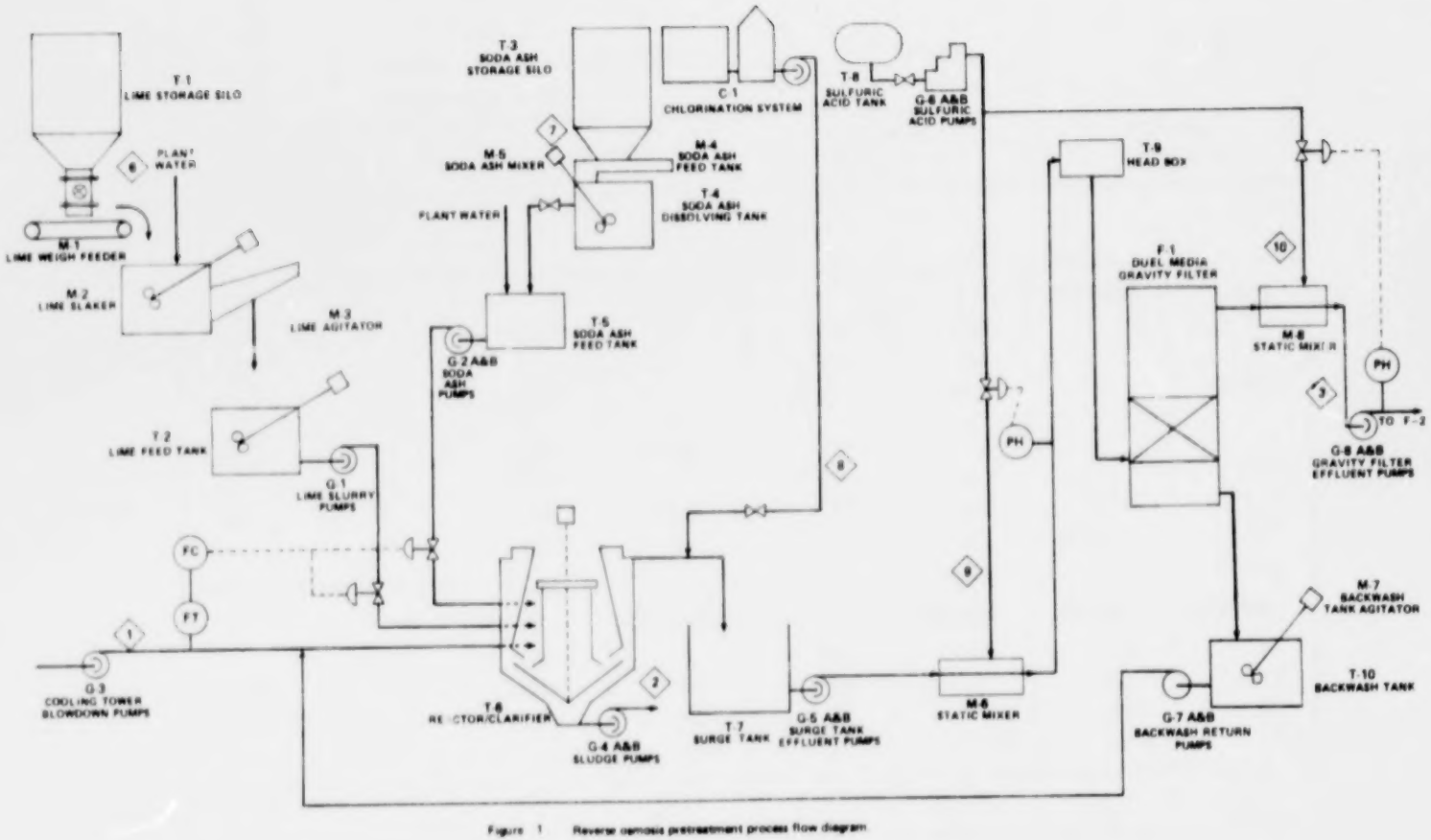


Figure 1 Reverse osmosis pretreatment process flow diagram

Figure 1a. Reverse osmosis pretreatment process flow diagram.

PARAMETER \ STREAM	WATER A ◇1	WATER A ◇2	WATER A ◇3	WATER B ◇1	WATER B ◇2	WATER B ◇3
pH	7.3	10.3	10.3	7.3	10.3	10.3
TEMPERATURE °F	80	80	80	80	80	80
FLOW GPM	1700	48	1700	1700	21	1700
M^3/H	386	11	386	386	5	386
LB/HR	849,660	23,990	849,660	849,660	10,496	849,660
SALINITY PPM AS ION	7522	–	7,619	5,562	–	5,650
Ca^{++}	400	32,650	50	533	38,517	50
Mg^{++}	400	5800	236	193	1,515	174
Na^{+}	1320	–	2,018	710	–	1,265
SiO_2	150	3975	37	5	–	5
Cl_2	–	–	–	–	–	–
$SO_4^{=}$	5000	–	5,027	3,820	–	3,847
HCO_3^{-}	44	–	43	35	–	–
Cl^{-}	208	–	208	266	–	43
SS	–	100,000	–	–	100,000	266

STREAM	CHEMICAL ADDITIVE	WATER A LB/HR	WATER B LB/HR
◇6	LIME 90% CaO	382	–
◇7	SODA ASH 98% Na_2CO_3	1,457	1,110
◇8	CHLORINE Cl_2	1.7	1.7
◇9	SULFURIC ACID 66° Be	16	17
◇10	SULFURIC ACID 66° Be	9	10

Figure 1b. Material balance for RO pretreatment process flow diagram.

All the pretreatment steps use equipment of proven commercial design. The reverse osmosis unit finally recovers 75 percent of the water content of the feed, rejecting the remaining 25 percent as a concentrated brine to storage ponds.

4.1.2 Operating Experience

Reverse osmosis is an established process for recovering purified water from saline feeds. During the past 10 years, it has gained wide acceptance for supplying potable water from brackish feed streams. More recently, reverse osmosis has been used to treat plating rinses, irrigation runoff, and municipal sewage. A measure of its acceptance is the recent award of contracts by the Bureau of Reclamation for membrane units to desalinate roughly 100 million gallons per day of agricultural drainage water. In addition to the above plants for the desalination of low salinity feeds (500 to 4,000 ppm of total dissolved solids), several membrane installations now produce potable water from seawater (35,000 ppm).

Figure 2 is a photograph of a portion of a large modern RO plant which produces potable water from seawater.

The performance of the membranes assumed in this report, both for recovery and salt rejection, is based on actual installations having feeds of similar composition. The pretreatment system is standard commercial practice for softening, clarifying, and removing silica from hard waters. It contains many of the features of the Yuma desalination plant being erected by the Bureau of Reclamation for the Colorado River Salinity Control Project.

4.1.3 Process Description

In the pretreatment process flow diagram (Figure 1), 1,700 gpm of cooling tower blowdown are delivered to the reactor-clarifier (RC) along with the solution of sodium carbonate (soda ash) and a lime slurry. The sludge formed in the reactor-clarifier is discharged to a storage pond (not shown) or to landfill. Alternatively, it may be thickened and filtered, and the filter cake sent to a kiln to recover the lime for reuse. However, this alternative is not included in the design study, since recovery of lime is not usually practical for the small amounts produced here.

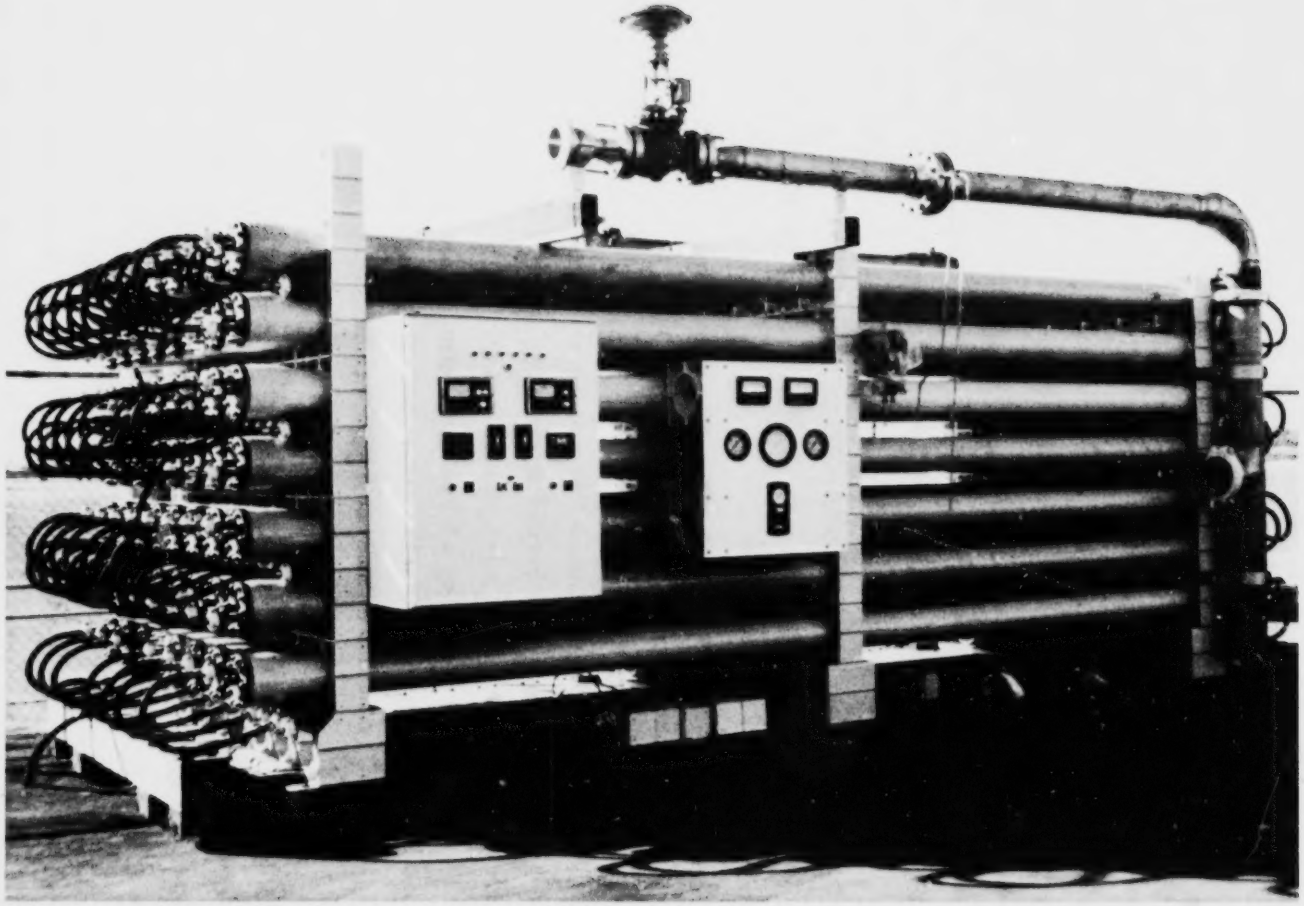


Figure 2. Photograph of portion of large seawater RO plant.
(Courtesy of Fluid Systems Division, UOP Inc.)

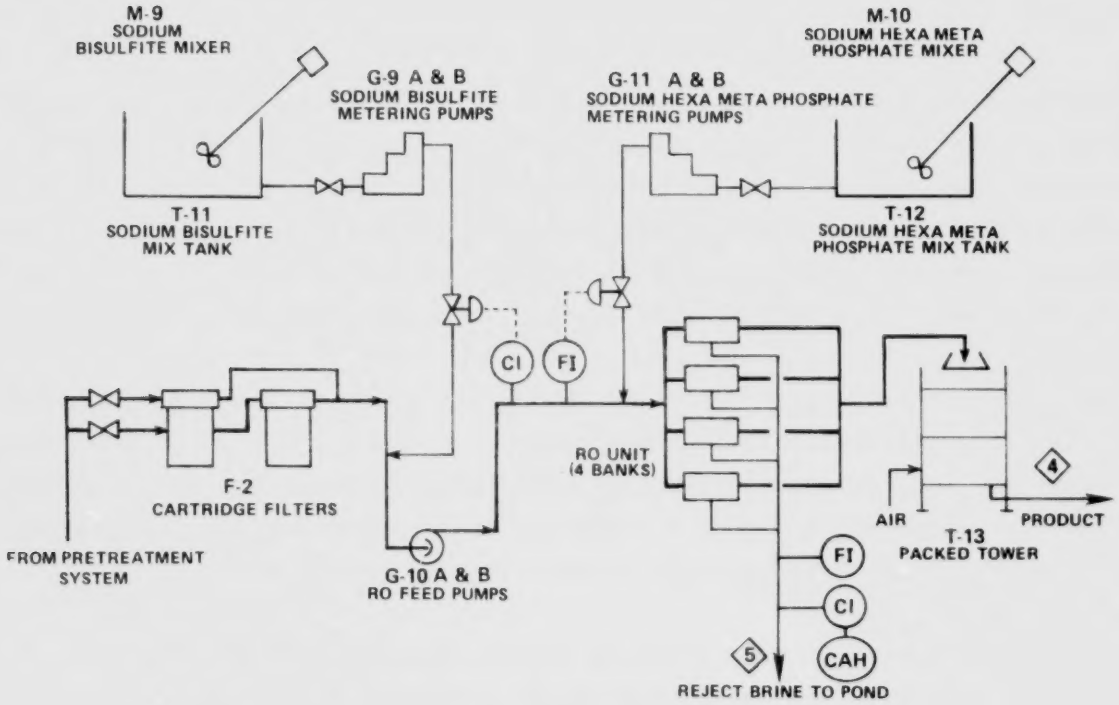
Chlorine is injected into the overflow from the RC to prevent growth of bacteria in the multi-media filter, the cartridge filters, and the lines downstream from the RC.

Because of the comparatively slow kinetics of the softening process, the downstream filter could become clogged as a result of post-precipitation. Consequently, the pH of the RC overflow is adjusted by adding acid. The acidified stream is then passed through a multi-media filter to reduce the suspended solids content. Periodic backwashing of the filter with part of the filtrate flushes the entrapped solids, and the flush water is recycled to the RC.

Additional acid is metered into the filtrate to reduce its Langelier Index to a value that will ensure freedom from calcium carbonate scale in the brine concentrate of the reverse osmosis unit and, thus, on the membranes themselves. (Refer to Appendix A.) The final polishing of the water is accomplished by passing it through 5-micron or 10-micron cartridge filters.

The pretreated water is then delivered to the RO system. (See Figure 3.) Before contact with the membranes of the RO units, the feed stream must be dechlorinated, either by injecting sodium bisulfite or by passing it through a bed of activated carbon. If polyamide membranes are used, residual chlorine must be reduced to zero. Cellulose acetate membranes, on the other hand, can tolerate up to 1.0 ppm of chlorine.

The pretreatment discussed above was aimed at maintaining a safe level of silica and calcium hardness in the RO reject stream. To maximize the recovery of purified water and thus achieve the greatest reduction in waste volume, it is common practice to operate the RO unit as close to the calcium sulfate scaling level as possible. In the present study, water recovery was assumed to be 75 percent. Under these conditions, any excursion or failure of the pretreatment system could result in serious calcium sulfate scaling of the membranes. Consequently, 5 ppm of sodium hexametaphosphate is injected to prevent scale formation by threshold inhibition, that is, to obstruct the growth of scale crystals.



QUANTITY \ STREAM		WATER A 4	WATER A 5	WATER B 4	WATER B 5
PH		5.2	6.3	5.2	6.3
TEMPERATURE	°F	80	80		80
FLOW	GPM	1275	425	1275	425
	M^3/H	290	97	290	97
	LB/HR	637,245	212,415	637,245	212,415
SALINITY PPM AS ION		475	28,855	384	21,358
Ca^{++}		4	187	0	200
Mg^{++}		17	893	0	700
Na^{+}		121	7708	126	4,696
SiO_2		8	125	1	17
Cl_2		0	0	0	0
$SO_4^{=}$		306	19083	214	14,638
HCO_3^{-}		4	73	7	151
Cl^{-}		15	786	36	956
SS		—	—	—	—

Figure 3. Reverse osmosis process flow diagram.

Finally, the volume of aqueous waste is reduced by passing it through the RO modules at 400 psig. Seventy-five percent of the feed is recovered in the form of relatively pure water containing 6 to 7 percent of the initial salt concentration. The balance of the feed, containing all the remaining solutes, is discarded to waste.

Details of the equipment and liquid streams are presented in Subsection 4.1.4.

4.1.4 Plant Design*

A pretreatment system is designed to supply water that meets the influent quality requirements of the selected RO system. Pretreatment consists of softening, pH adjustment, chlorination, filtration, sodium bisulfite addition, and sodium hexametaphosphate (SHMP) addition. Each pumping system is provided with an installed spare pump to satisfy the high reliability requirements of the treatment train. All pumps are sized at 110 percent of required capacity.

The pretreatment system for RO is sensitive to the composition of the cooling tower blowdown feed stream, so Water A and Water B are considered separately. Figure 1 shows the pretreatment system. Water A (see Table 2) is softened by the cold lime soda process. Lime and soda are mixed with cooling tower blowdown in the reaction zone of a 60 foot diameter reactor-clarifier (RC) of steel construction. An 1,870 gpm pump transfers the blowdown from the cooling tower basin to the RC. A lime system supplies lime to the RC as a 10 percent slurry. This lime system consists of a 140 ton capacity silo, which stores a 30 day supply of chemical lime (90 percent CaO), a 380 pound per hour gravimetric feeder, and a slaker with a capacity of 380 pounds per hour. The lime slaker discharges a slurry to a 2,600 gallon capacity feed tank which provides 4 hours' retention time. A one horsepower agitator maintains the feed tank solids in suspension,

*In the specification and purchase of water treatment equipment, English units are commonly used for dimensions, fluid flows, and energy transfer and consumption. The reader is referred to Appendix B for the conversion table to SI units.

and a 10 gpm pump supplies the lime slurry from the feed tank to the RC. A soda ash system delivers soda ash to the RC as a 10 percent solution.

A 525 ton capacity silo stores a 30 day supply of soda ash. A gravimetric feeder supplies soda ash at 1,450 pounds per hour from the silo to a dissolving tank with 1,270 gallons capacity and 30 minutes' retention time. A one horsepower agitator is mounted in the dissolving tank to dissolve the soda ash in water. The soda ash solution flows by gravity to a feed tank with 8,500 gallons' capacity and 4 hours' retention time. A 30 gpm pump delivers the soda ash solution from the feed tank to the RC. Instrumentation is provided for automatic adjustment of the lime and soda ash feed rates based on the cooling tower blowdown flow rate.

As the blowdown, lime, and soda ash react in the RC, solids form. These solids eventually settle as a sludge and flow to a sump. A 50 gpm pump discharges the sludge from the sump to receiving facilities for final disposal at a landfill. Clarified, softened water flows over the RC weir into a 28,800 gallon capacity surge tank with 15 minutes' retention time. A chlorination system injects approximately 40 pounds of chlorine per day to the surge tank influent. This results in a chlorine concentration of approximately 2 ppm in the surge tank contents.

An 1,870 gpm pump transfers water from the surge tank to a 19,800 gallon capacity head box with 10 minutes' retention time. Water flows by gravity from the head box to an upflow dual media filter. The head box outlet is situated above the filtered water effluent to provide sufficient head for the water to flow upward through the filter media and into the filtered water storage tank. The filter sizing is based on a loading rate of 3 gpm per square foot. The filter is 28 feet in diameter and is divided into three compartments of equal area. Each compartment of the filter is backwashed automatically when the pressure drop across the compartment unit reaches a predetermined level. Under design conditions, it is anticipated that each compartment will be backwashed twice a day. The backwash cycle lasts 4 to 5 minutes and the backwash rate is 15 gpm per square foot. A 16,000 gallon backwash holding tank is sized to accommodate the

total flow from the backwash cycle of one compartment. A 10 horsepower agitator maintains the solids in suspension in the backwash tank. A 60 gpm pump continuously recycles the backwash water to the RC.

A sulfuric acid system is provided to add 66°Be sulfuric acid to the system. A 1,700 gallon tank stores a 30 day supply of acid. A metering pump with a range from 1 to 5 gph delivers acid at approximately 24 gallons per day (gpd) to the surge tank effluent and 14 gpd to the gravity filter effluent. An 8 inch diameter, 1 foot long, in-line static mixer is provided at each acid injection point to ensure adequate mixing of acid and water.

Figure 3 shows the RO system. An 1,870 gpm pump is used to pump the gravity filter effluent through one of the two full-sized cartridge filters which are installed in parallel. When one filter is being backwashed, the flow is diverted to the standby filter.

A metering pump with a discharge capacity ranging from 1 to 10 gph adds sodium bisulfite as a 10 percent solution to the cartridge filter effluent. A 1/4 horsepower mixer dissolves one day's requirement of sodium bisulfite (82 pounds as sodium metabisulfite) in water in a 200 gallon tank. The discharge from the metering pump is regulated by the residual chlorine concentration in the RO feed pump discharge.

The cartridge filter effluent is pumped by an 1,870 gpm, 400 psi, 625 hp RO feed pump through the RO system. Immediately upstream of the RO system, sodium hexametaphosphate (SHMP) is added as a 10 percent solution to the RO feed pump discharge by a metering pump with a 2 to 10 gph discharge range. A 1/4 horsepower mixer dissolves one day's supply (100 pounds) of SHMP in water in a 260 gallon tank. The SHMP feed rate is proportioned to the RO feed pump discharge rate.

The RO system consists of membranes arrayed in four banks of modules (individual vessels containing the membranes) in a three-stage configuration. The first stage of each bank has 19 modules, the second stage has 8 modules, and

the third stage has 6 modules. Each bundle of membranes is enclosed in an epoxy-coated steel cartridge 10-3/4 inches in diameter and 48 inches long. Seventy-five percent of the pretreated water applied to the RO system is recovered as product, and the minimum salt rejection is 90 percent. Any hydrogen sulfide and carbon dioxide present are removed from the permeate by forced-draft degasification in an 8 foot diameter by 16 foot high packed tower. The packed tower is constructed from carbon steel with coal tar epoxy coating. The permeate contains 475 ppm of total dissolved solids and flows at 1,275 gpm. The reject stream contains 28,855 ppm of total dissolved solids and flows to solar evaporation ponds at 425 gpm. The ponds provide an evaporative surface area of 160 acres and are sized on the basis of 4.0 feet of assumed net evaporation per year.(Ref. 1) The compositions of the permeate and the reject stream appear in Table 4.

TABLE 4. WATER A: RO PERMEATE AND REJECT

	Composition Expressed as ppm of Ion	
	Permeate	Reject
Na	121	7,708
Ca	4	187
Mg	17	893
Cl	15	786
SO_4	306	19,083
HCO_3	4	73
SiO_2	8	125
Sum of ions	475	28,855
pH	5.2	6.3
T	80°F	80°F

A half-acre emergency pond is provided to store three days' feed to one of the four RO banks to allow for complete shutdown and repair of one bank.

The pretreatment system for Water B differs from that for Water A in that lime is not required in the softening step, and the soda ash requirement is reduced from 17.5 tons per day to 13.3 tons per day as 98 percent sodium carbonate. Table 5 shows the soda ash and sludge handling equipment capacities for Waters A and B where these values are not identical.

TABLE 5. SODA ASH AND SLUDGE HANDLING EQUIPMENT FOR WATERS A AND B

Equipment	Water A	Water B
Soda ash silo	17,180 ft^3	14,730 ft^3
Soda ash feeder	1,450 lb/hr	11,000 lb/hr
Soda ash feed tank	1,140 ft^3	863 ft^3
Soda ash feed pump	30 gpm, 1 hp	25 gpm, 0.5 hp
Sludge pump	30 gpm, 1 hp	25 gpm, 0.5 hp

The RO system required to treat Water A has 19 modules in the first stage of each bank, while the RO system for Water B has 18 modules in the first stage of each bank. The two RO systems are otherwise identical, with 75 percent overall system recovery and 90 percent minimum salt rejection in each case.

The permeate contains 384 ppm of total dissolved solids and flows at 1,275 gpm. The reject stream contains 21,358 ppm of total dissolved solids and flows to the solar evaporation ponds at 425 gpm. The evaporative surface area required is the same for Water A and Water B. Table 6 shows the compositions of the permeate and reject stream when Water B is treated by RO.

Table 7 presents the design criteria for the plant equipment. Figure 4 shows the layout of the RO plant.

TABLE 6. WATER B: RO PERMEATE AND REJECT

	Composition Expressed as ppm of Ion	
	Permeate	Reject
Na	126	4,696
Ca	0	200
Mg	0	700
Cl	36	956
SO_4	214	14,638
HCO_3	7	151
SiO_2	1	17
Sum of ions	384	21,358
pH	5.2	6.3
T	80°F	80°F

TABLE 7. SUMMARY OF DESIGN CRITERIA FOR RO OPTION

Equipment	Design Criteria
Bulk storage for lime, soda ash, and sulfuric acid	30 day supply
Feed tanks	4 hours' retention time
Lime weigh feeder	380 lb/hr capacity
Chemical lime slaker	380 lb/hr capacity
Soda ash weigh feeder	1,450 lb/hr capacity
Soda ash dissolving tank	30 minutes' retention time
Chlorine system	2 ppm chlorine dosage
Reactor-clarifier	Surface rate of 0.75 gpm/ft^2 with 20 percent additional area
Surge tank	15 minutes' retention time
Head box	10 minutes' retention time
Dual media gravity filter	Surface rate of 3 gpm/ft^2
Cartridge filters	2 full size systems
Sodium bisulfite mix tank	1 day's supply
Sodium hexametaphosphate mix tank	1 day's supply
RO system	4 banks, each with 480,000 gpd capacity
Packed column	Loading rate of 35 gpm/ft^2, packed height 8'0"

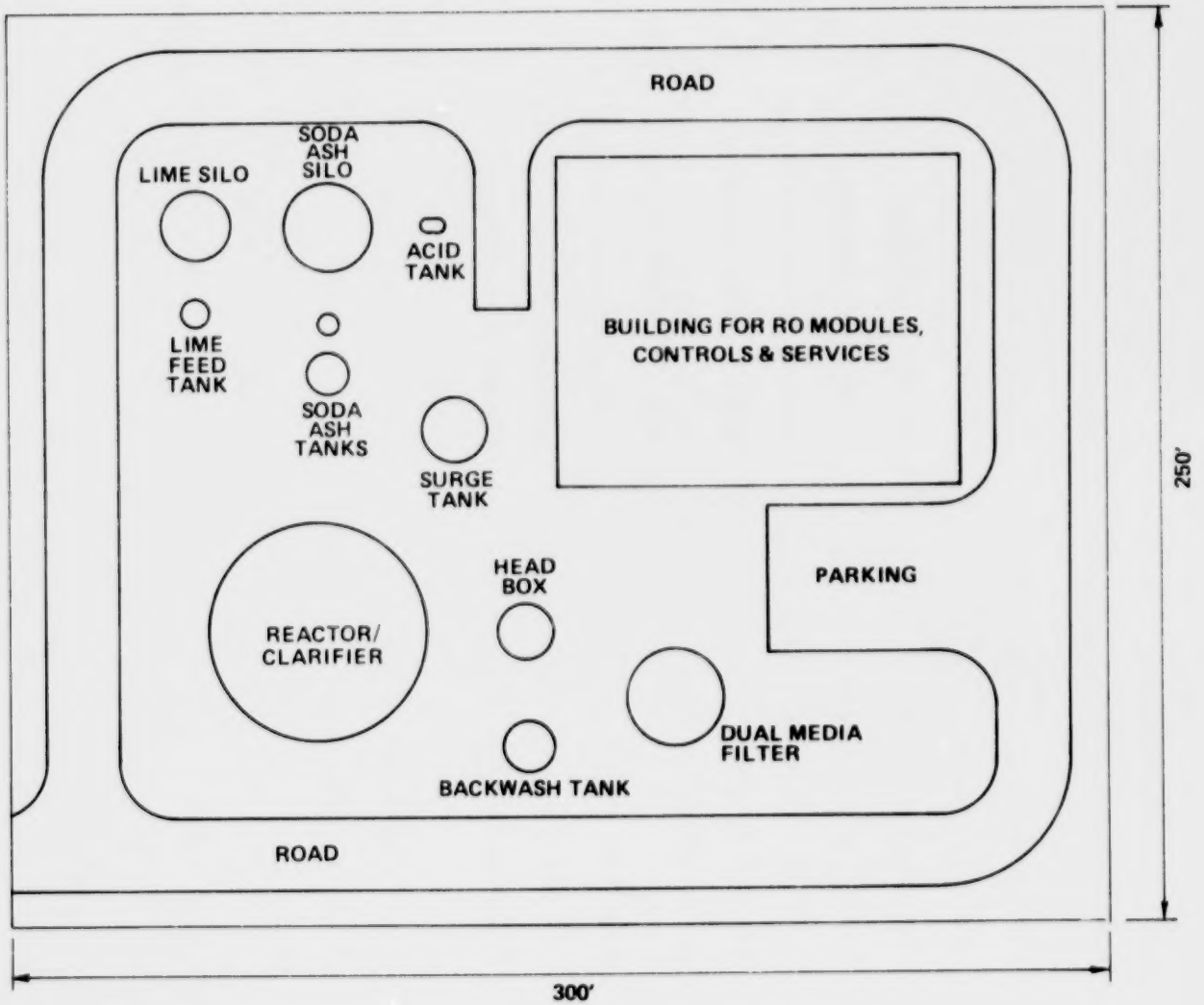


Figure 4. Plot plan for RO option.

4.1.5 Process Economics

The investment in the pretreatment plant is based on quoted prices for the major equipment components. For the RO unit, quotations were solicited on the system as a complete package, including pumps, instruments, and controls. Prices were quoted by three vendors, differing by only a few percent. The average of these quotations was used as the cost basis. Factors based on experience in the costing of water treatment plants were applied to cover site preparation, foundations, installation costs, contractors' fees, and contingencies. Pond requirements were calculated on the basis of an assumed 30 year useful life. The total installed plant cost, including ponding for the storage of blowdown, was estimated at $13,900,000.

Treatment costs include operating costs, plus a capital charge based on a 30 year plant life. Membrane replacement charges assume a membrane life of three years, in agreement with current experience in brackish water plants incorporating a thorough pretreatment of the feed. The calculated energy consumption is 5.3 million kWh per year, primarily for pumping power.

Three parameters were investigated.

- The costs of lined ponds were assumed to be $40,000, $50,000, and $60,000 per acre, respectively
- Costs were developed for power at 2¢, 4¢, and 6¢ per kWh
- The fixed charge rate was assumed at 12.5 percent and 16 percent, respectively

The results of the cost analysis are presented in Figures 5 and 6. Figure 5 indicates a comparatively minor dependence of RO treatment cost on the price of energy. This is because the large ponding requirements of RO place emphasis on plant investment and resulting capital costs. In the cases studied here, capital costs comprise 57 to 68 percent of the total water treatment cost. Energy, on the other hand, contributes from 4 percent to 11 percent, the actual value depending on the fixed charge rate and kWh prices assumed.

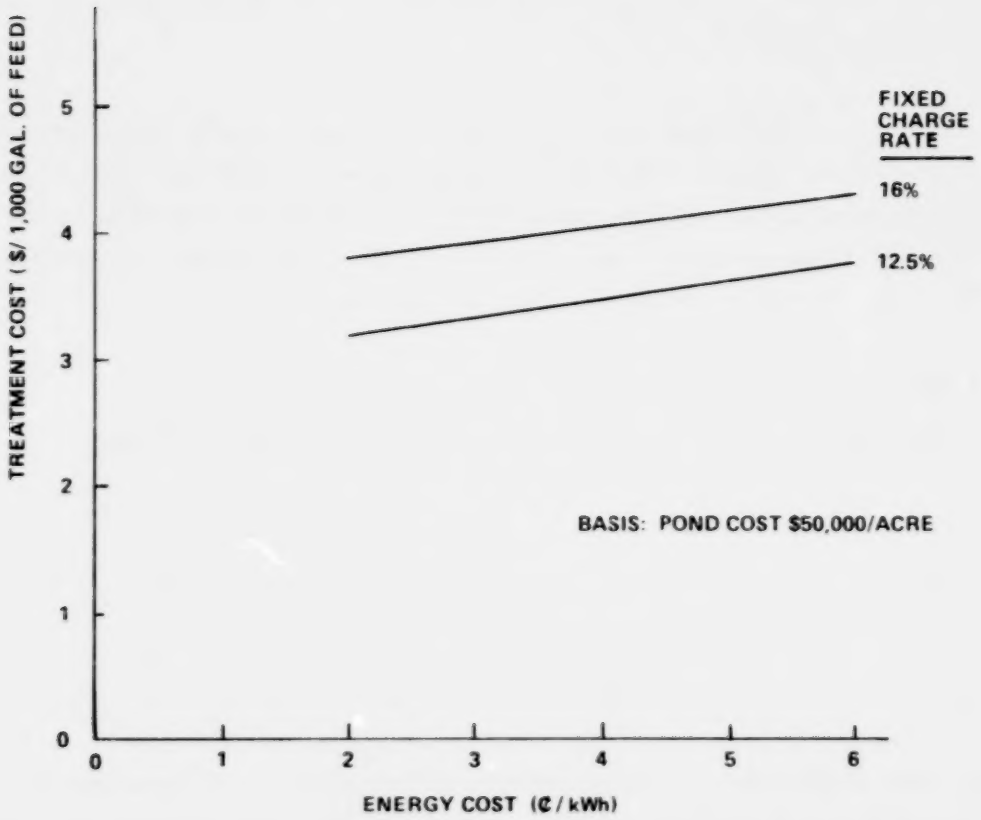


Figure 5. Effect of energy cost on RO treatment costs.

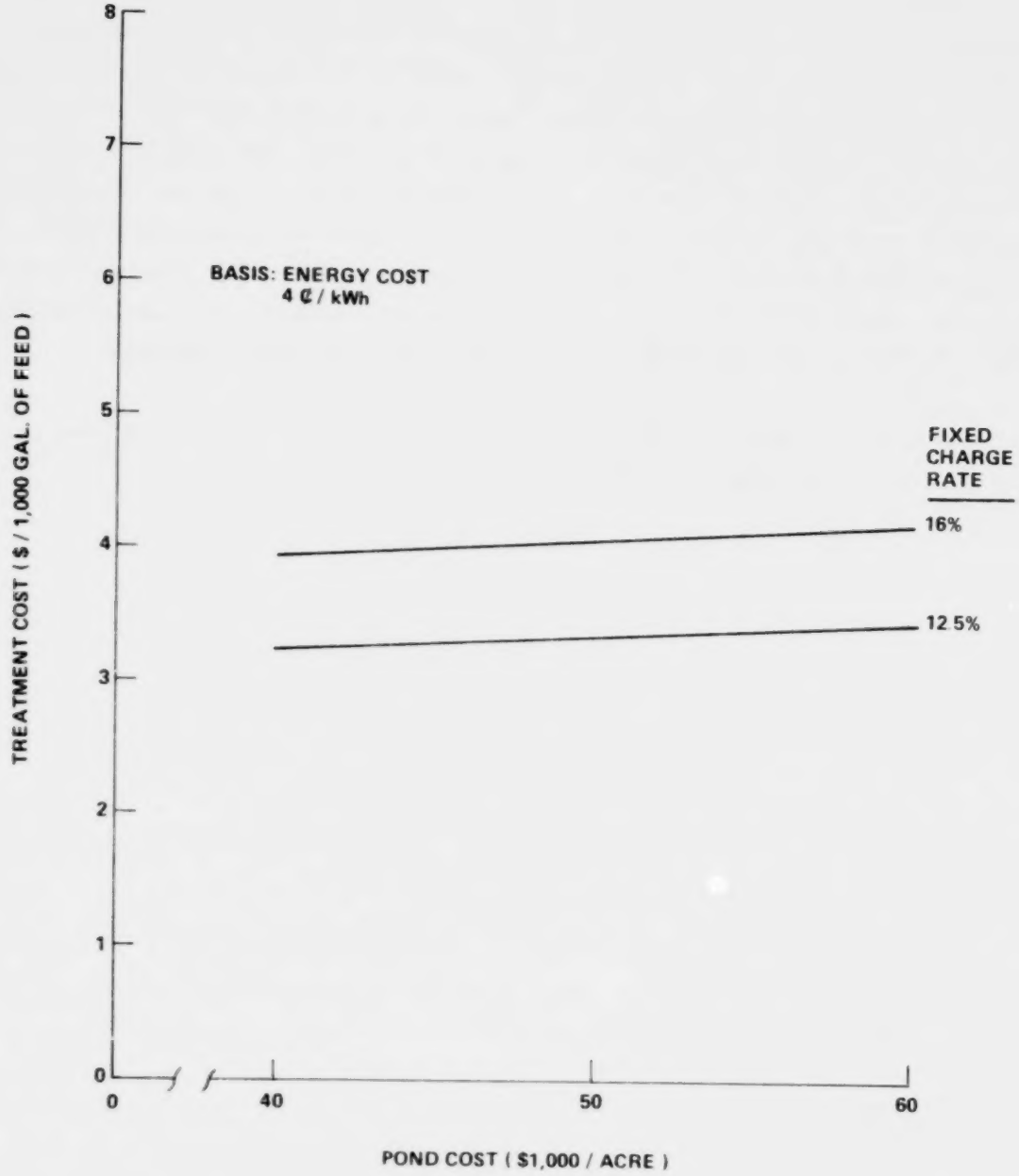


Figure 6. Effect of pond cost on RO treatment costs.

The comparative importance of assumed pond costs is apparent in Figure 6, where an increase from $40,000 to $60,000 per acre increases the treatment cost by roughly 20 percent. It is possible to increase the recovery of the RO plant from the 75 percent value assumed in this study to 85 percent or possibly higher for the particular feed waters assumed here. As a result, the flow of brine to the ponds by the RO plant would decrease by 36 percent. The size of the ponds and their cost would decrease by a lesser amount, since a fraction of the pond volume is required for the crystals precipitated from the evaporating brine. The increased recovery of the RO plant, however, would require a more thorough and more costly pretreatment of the cooling tower blowdown. The choice of the exact process conditions would require a more detailed cost comparison.

For comparison of costs of the three candidate processes analyzed in this report, please see Subsection 4.5.

4.2 VAPOR COMPRESSION EVAPORATION

4.2.1 Design Basis

In contrast to reverse osmosis, the Resources Conservation Company's vapor compression evaporator investigated in this study can tolerate the presence of scale formers. Deposition on the plant components is prevented by maintaining a comparatively high concentration of calcium sulfate crystals in the brine to provide nuclei on which the scale will deposit in preference to the equipment and piping. If any scale deposits on the evaporator tube surfaces, it is scoured off by the recirculating slurry.

The seed slurry is not present, however, in the feed preheater. (See the process description in Subsection 4.2.3.) There the concern is to avoid calcium carbonate scale, which tends to deposit on heated surfaces because of its inverse temperature solubility characteristics. To prevent calcium carbonate deposition, the pH of the cooling tower blowdown is adjusted to a negative Langelier Index by adding acid.

The feed assumed for the vapor compression plant is unsoftened cooling tower blowdown Water A, having the composition shown in the first column of Table 2.

4.2.2 Operating Experience

Vapor compression evaporation is an old and established technique that has been used principally for the recovery of potable water from seawater. Its use for concentrating cooling tower blowdown was pioneered within the past decade by Resources Conservation Company. Figure 7 is a photograph of a commercial VC plant.

At present, ten units, ranging in feed capacity from 10 to 700 gpm, are in operation in waste concentration service, mostly in power generating stations. Seven more are under construction. The evaporators recover up to 98 percent of the water in the feed in the form of a very pure product (2 to 6 micromho conductivity, corresponding roughly to 1 to 3 ppm of total dissolved solids).

Figure 7. Photograph of a commercial VC plant.
(Courtesy of Resources Conservation Company)

The product can be processed in a mixed-bed ion exchange polisher to yield ultrapure water for boiler makeup.

The on-stream factor of all the plants investigated has been very high.

4.2.3 Process Description

The vapor compression evaporator (VCE) studied in this report is fundamentally a vertical tube evaporator in which the liquid being evaporated descends as a thin film on the inside of tubes arranged in a bundle within an evaporator shell. Steam within the shell condenses on the exterior walls of the tubes, thereby giving up its latent heat to evaporate a fraction of the water content from the liquid film in the tubes. By returning the partially concentrated fluid to the upper plenum (water box), further evaporation is achieved to produce a fairly concentrated brine.

The VCE differs from a conventional vertical tube evaporator in its lack of an external condenser for the steam generated during the evaporation process. Instead, the steam is compressed to raise its temperature slightly (usually $5^{\circ}F$ to $12^{\circ}F$) and is then delivered to the shell of the heat exchanger section where it condenses on the tubes as described in the preceding paragraph. That is, the steam chest itself acts as a condenser. The source of evaporative energy is the mechanical energy input of the compressor rather than the thermal energy of a steam supply.

Referring to the schematic diagram in Figure 8, the cooling tower blowdown (pH adjusted by acid addition to avoid calcium carbonate scaling of the heat exchanger) is pumped through a plate-and-frame type heat exchanger, where it is preheated by heat exchange with product water. Air and carbon dioxide are stripped from the feed stream in a vacuum deaerator.

Deaerated brine is distributed over the upper tubesheet of the evaporator and flows down as a thin film on the inside wall of each of the tubes. Hot vapor condensing on the outer wall of the tubes contributes the heat required to

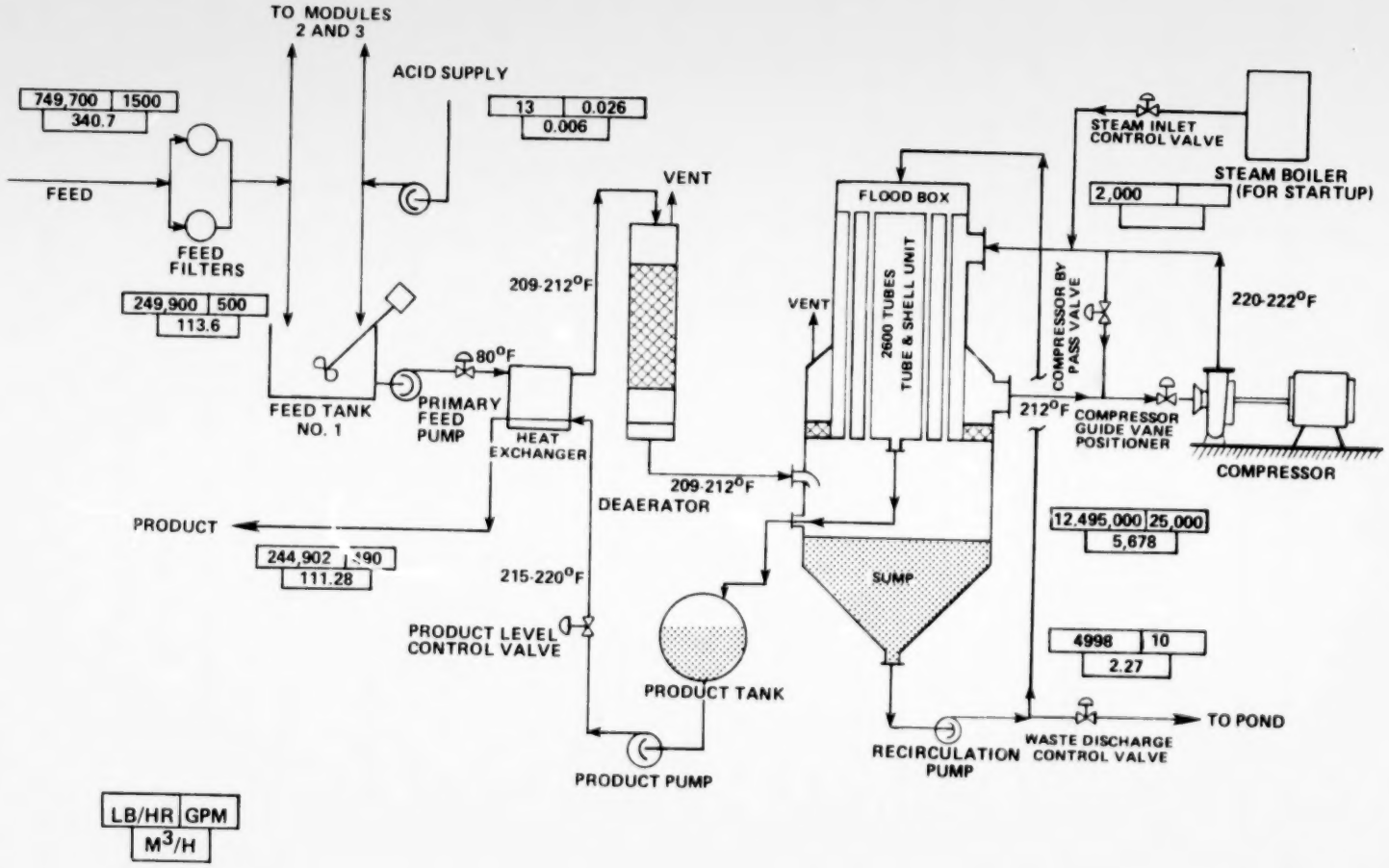


Figure 8. Vapor compression evaporation module.

evaporate a portion of the water contained in the feed. The brine that falls to the bottom of the evaporator is pumped to the top to repeat its descent through the tubes, thereby evaporating an additional fraction of its water content. A portion of the recirculating brine, together with any precipitates suspended in it, is bled from the loop and is discharged to waste. Since it represents only a small fraction of the volume of feed to the evaporator, no attempt is made to recover its sensible heat.

The vapor generated within the tubes is discharged through a mist eliminator and is withdrawn by the vapor compressor and delivered to the shell side of the evaporator; there it condenses on the outside of the tubes as described in the preceding paragraph. The compressor serves a double purpose:

- It contributes the energy required for the evaporative process
- By the adiabatic compression of the vapor, it raises its temperature to provide the thermal driving force across the wall of the tubes

The small steam boiler supplied at no additional cost by the system manufacturer (see Figure 8) is used for plant startup only.

The condensate is withdrawn from the evaporator shell and cooled by passage through the heat exchanger, where it preheats the feed as previously described. The purity of this condensate is monitored by a conductivity cell in the product water line.

The vapor compression evaporator cycle is very attractive because of its high energy effectiveness under commercially attainable operating conditions. Energy consumption for distillation as low as 70 kWh per 1,000 gallons of cooling tower blowdown fed to the evaporator has been achieved in operating, full-size plants. In equivalent heat consumption this corresponds, for example, to a multistage flash distillation plant operating at the very high economy ratio of 12 pounds of evaporation per 1,000 Btu's supplied, compared with the more common economy ratio of 8. The overall energy consumption, including the power to drive the pumps, will range from 70 to 90 kWh per 1,000 gallons.

There are several features inherent in a VCE that strongly affect its economics. All other things being equal, the smaller the temperature rise and, hence, the compression ratio, the higher the overall energy efficiency of the process. On the other hand, a small temperature rise requires a large heat transfer area. Consequently, the choice of operating conditions is a trade-off between cost of energy consumed and plant investment. A second consideration concerns the operating temperature of the VCE. The choice of a low operating temperature will greatly decrease corrosion but will produce vapor of high specific volume, thereby increasing the size of the compressor and vapor lines. The system selected for analysis in this report operates at approximately 212°F.

To provide the initial charge of vapor to the compressor, a small auxiliary boiler supplies steam to the evaporator for a short time at startup. In addition, stable operation is not achieved for the first few hours after an extended shutdown to permit the scale crystals, suspended in the brine, to revert to the optimum crystal form for seeding subsequent scale deposits.

4.2.4 Plant Design

The raw cooling tower blowdown requires only minimal pretreatment before it enters the vapor compression evaporator system. The pretreatment consists of filtration and pH adjustment.

Figure 8 shows one of the three vapor compression evaporation modules. Blowdown from the cooling system is discharged to the concentrator at 1500 gpm. Two 316 stainless steel 1,650 gpm pumps (one an installed spare) are provided to transfer the blowdown from the cooling tower sump through two on-line filters to three feed tanks. Two spare filters are provided to enable continuous operation of the evaporator during the backwash cycle. Each feed tank is constructed from fiberglass reinforced plastic and has a capacity of 6,460 gallons to provide 10 minutes' retention time for one-third of the total cooling tower blowdown flow. A sulfuric acid system adds about 12 gpd of 66°Be sulfuric acid to each feed tank to lower the pH of the blowdown from 7.3 to within the range of 5.5 to 6.0. A one hp agitator provides adequate mixing of the acid and water in each tank.

The evaporator system consists of three modules, each with 500 gpm capacity. Each feed tank supplies water to a 316 stainless steel 550 gpm feed pump which, in turn, feeds one module. Each module consists of a heat exchanger, a deaerator, an evaporator with a brine pump, a recirculating pump, a product tank, a product pump, and a compressor. The feed pump discharge enters the heat exchanger which brings the temperature of the feed to near its boiling point by recovering the sensible heat contained in the hot product (condensate) stream from the evaporator. The heat exchanger is of a plate-and-frame type with titanium plates. The approach of the heat exchanger is $6^{o}F$.

The heated stream next enters a counterflow, packed column, atmospheric deaerator measuring 18 inches in diameter and 4 feet high. The deaerator has a 316L stainless steel shell and is packed with plastic Berl saddles. Carbon dioxide, nitrogen, and oxygen are removed in the deaerator and vented to the atmosphere. The feed leaves the deaerator at $209^{o}F$ to $212^{o}F$ and flows by gravity to the evaporator sump. Here, it mixes with the concentrated slurry and is continuously recirculated by a 25,000 gpm pump. The evaporator is 12 feet in diameter and is 80 feet high. It has a 316L stainless steel shell and a tube bundle consisting of 2,600 titanium tubes, each having an outside diameter of 2 inches and a thickness of 0.028 inch. The tubes are 50 feet long, and provide 64,000 square feet of heat transfer surface.

The recirculated flow is distributed to the inside wall of each tube as a thin film. Water is evaporated as the film falls down inside the tubes. The steam formed passes through a mist eliminator and enters the suction line of the compressor at approximately $209^{o}F$ to $212^{o}F$. The compressor has a 3,500 hp motor and compresses the steam by 2 psi to raise its condensation temperature to $220^{o}F$ to $222^{o}F$, about $6^{o}F$ above the boiling point of the recirculating brine. As the steam condenses on the shell side of the tubes, it gives up its heat of condensation and is collected in a 316L stainless steel product tank. The latent heat of condensation provides the energy required to evaporate the brine inside the tubes.

The brine in the sump contains 301,700 ppm total dissolved solids and 82,000 ppm suspended solids. The product contains less than 10 ppm total dissolved solids. The concentration factor of the system is 50.

The hot (220°F to 222°F) product is pumped back through the heat exchanger by a 316L stainless steel 75 psi pump. The product stream is about 98 percent of the feed stream. A device senses the salinity of the brine in the sump continually, and maintains the sump total dissolved solids concentration in the desired range by controlling a waste flow valve. The waste flow, normally about 2 percent of the feed stream, is discharged to a 50 acre solar evaporation pond. The pond size is based on a net evaporation rate of 4.0 feet per year (Ref. 1) and is surrounded by a dike with a 2:1 slope. The water depth in the pond is 10 feet.

At startup, each unit is seeded with calcium sulfate crystals to produce a slurry. Twelve hours are required to heat the slurry up to the desired temperature. The slurry is then "heat soaked" for 24 hours. A boiler is provided to supply 2,000 pounds of steam per hour at 15 psi for startup from a cold start.

The service water requirement for the evaporator system is 60 gpm. The air requirement for the controllers is 100 cfm at 3 psi. The evaporator system includes all controls for automatic sensing and recording of flow rates, temperatures, and pressures.

A 0.85 acre emergency pond is provided to store three days' flow to one of the three modules. The pond is lined with Hypalon and has a water depth of 10 feet. An earthen dike with a 2:1 slope provides 2 feet freeboard. The layout of the vapor compression evaporation plant is shown in Figure 9.

4.2.5 Process Economics

The plant investment for vapor compression evaporation is based on the turnkey price of an analogous system recently installed for concentrating cooling tower blowdown. Adjustments were made to include site preparation, roads, fences, and service lines to the plant. The total installed plant cost, including ponds for the storage of blowdown, was estimated at $15,200,000.

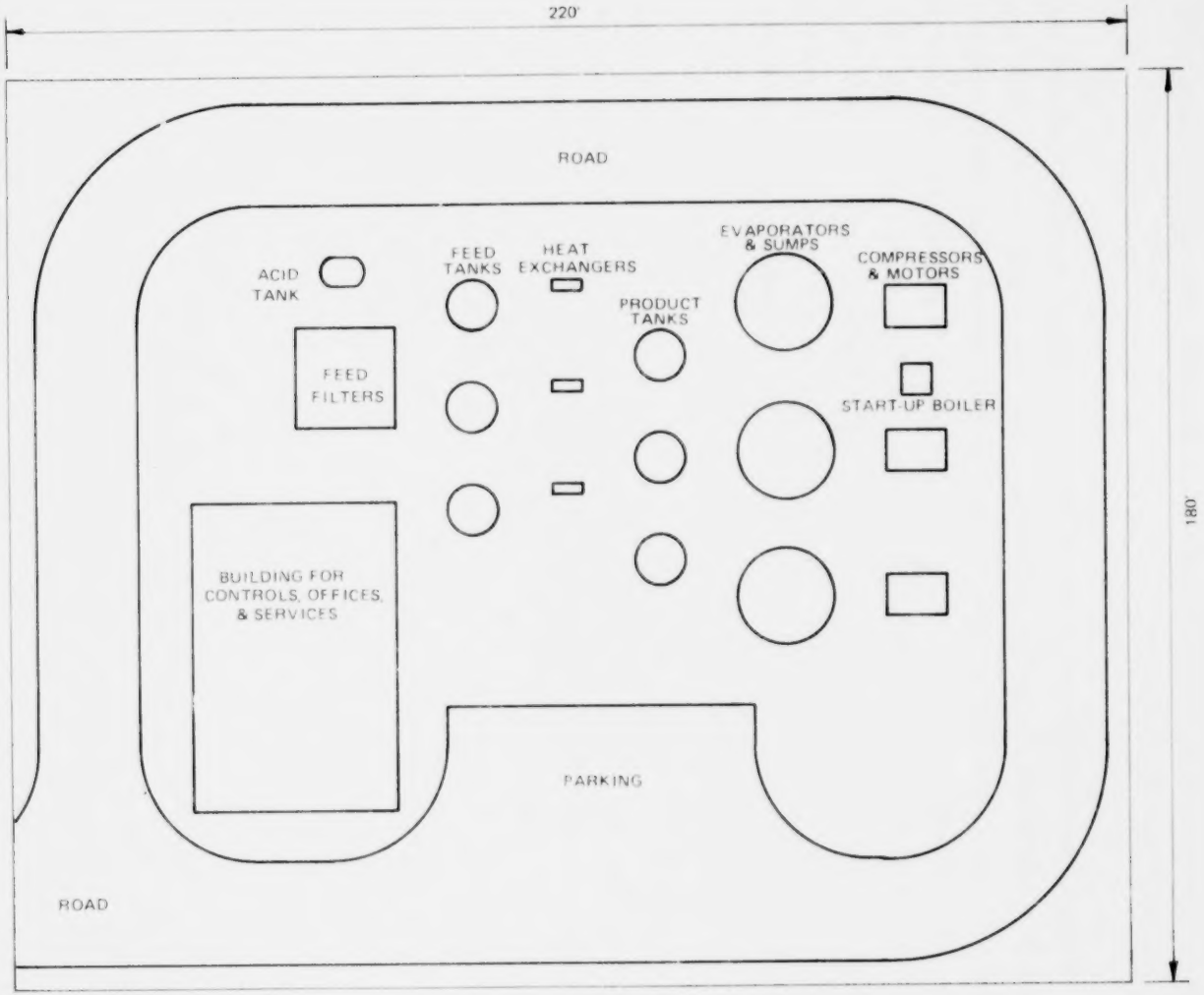


Figure 9. Plot plan for vapor compression evaporation option.

The energy consumption was assumed to be 81 kWh per 1,000 gallons of feed, roughly the median value for vapor compression plants treating similar feed streams. In actual practice, the supplier guarantees the energy consumption for each plant quoted.

Vapor compression evaporation is one of the least energy intensive of all the distillation processes (provided, of course, that the energy for the evaporation processes is not "free" as has been assumed in the case of the VTFE in this report). Nevertheless, the cost of power represents a substantial fraction of the overall cost of treating the cooling tower blowdown by this process. In the cases analyzed here, energy consumption contributes 37 to 63 percent to the overall treatment cost, the actual value varying with the fixed charge rate and kWh prices assumed. The importance of energy cost is emphasized by the steep rise in the treatment curves in Figure 10.

In contrast to the 25 percent brine rejection of RO, the VC plant rejects only 2 percent of the feed stream, recovering the remaining 98 percent as purified water. Consequently, Figure 11 shows a very small dependence of treatment costs on the cost of ponds.

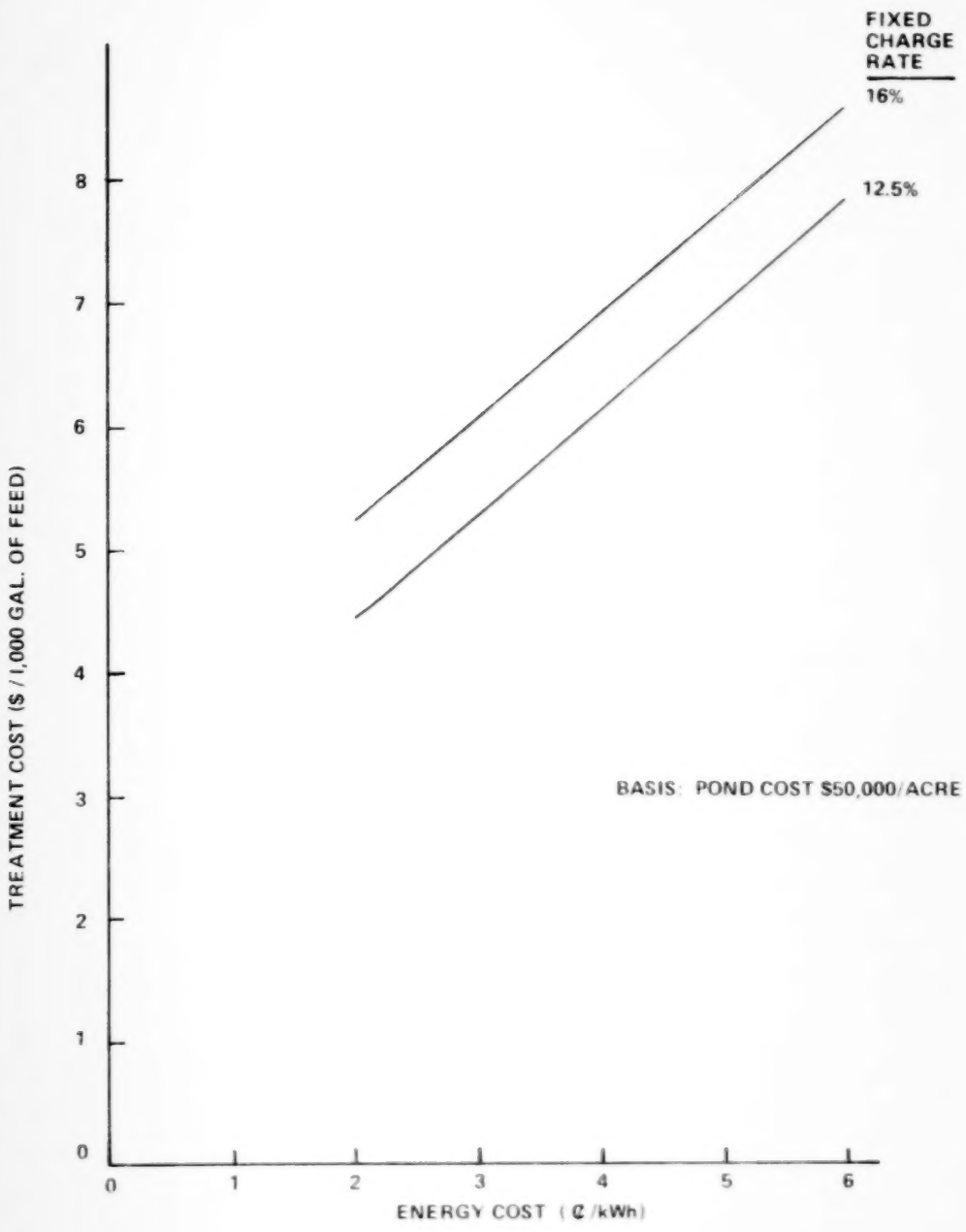


Figure 10. Effect of energy cost on VC treatment costs.

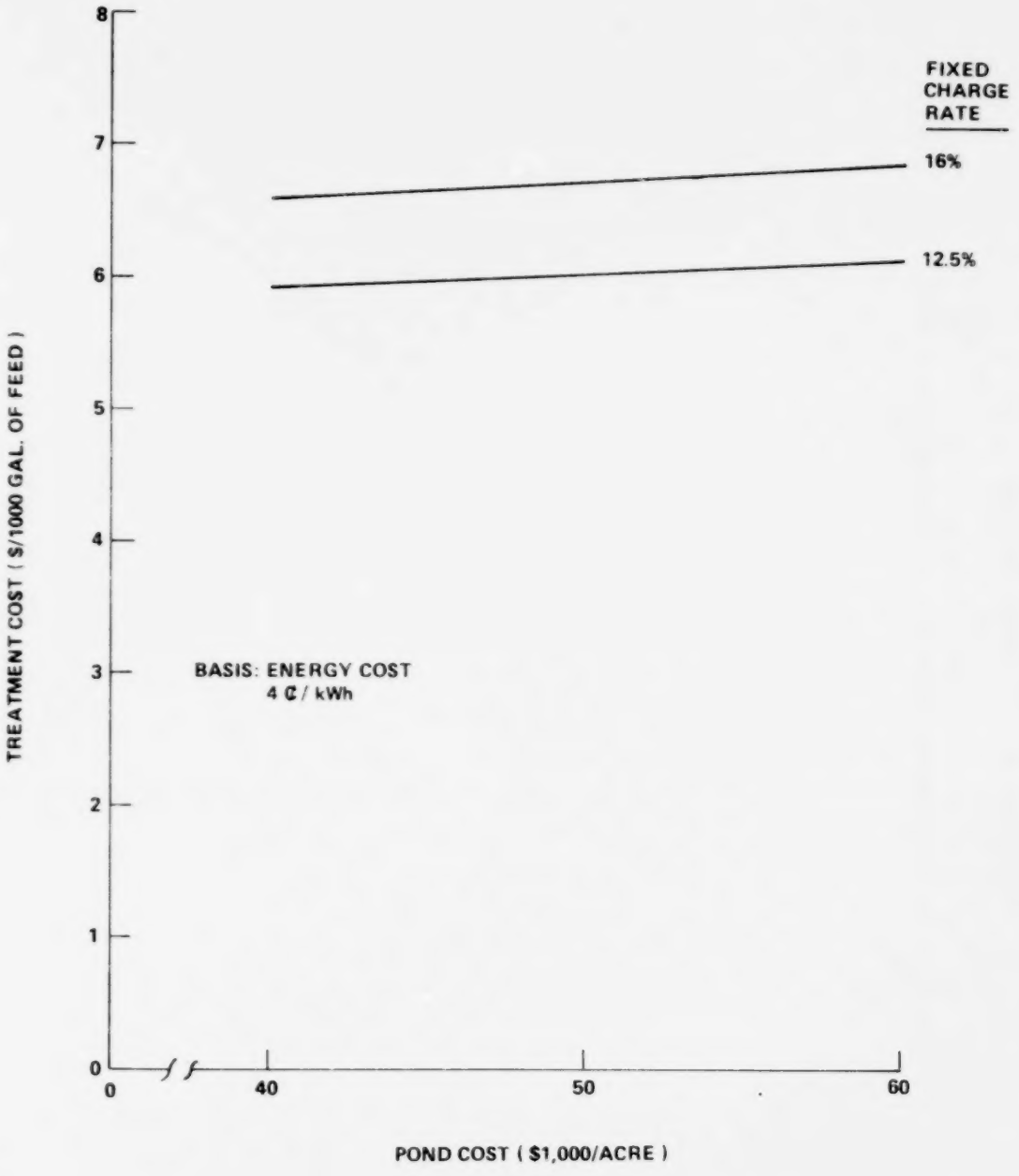


Figure 11. Effect of pond cost on VC treatment costs.

4.3 VERTICAL TUBE FOAMING EVAPORATION

4.3.1 Design Basis

This section is devoted to discussion of vertical tube foaming evaporation (VTFE). At the time of writing of this report, however, no VTFE units are in commercial operation, nor have any test data been generated by the recently constructed pilot plant. Consequently, the technical and economic analysis is based on the vertical tube evaporator (VTE) rather than the VTFE, as agreed by the EPA. Because of the lack of data, a very conservative approach has been adopted. If the pilot plant substantiates the encouraging results previously reported by the Sea Water Conversion Laboratory of the University of California for tests on the earlier small pilot unit, a reduction in the size and plant investment reported in this section may be anticipated.

The vertical tube foaming evaporator (VTFE) system consists basically of a conventional vertical tube, recirculating type evaporator to achieve the major portion of cooling tower blowdown concentration. It is followed by a similar but smaller evaporator for carrying the concentration to the crystallizing stage. The novel feature is the addition of a small amount of surfactant to the feed. The surfactant has been shown to provide three advantages:

- The rate of deposition of scale formers on the heat transfer surfaces and other plant components is drastically reduced
- A foaming surfactant produces a comparatively stable two-phase fluid in the evaporator tubes; this lowers hydrostatic losses enought to permit recirculation of brine in the upflow mode without a pump when operating at somewhat elevated temperatures (above 150°F)
- The overall heat transfer coefficient in the tubes is improved by adding the surfactant.

Because this process can tolerate the presence of certain scale formers, either of the two cooling tower blowdown compositions would be suitable as feed to the VTFE. For this study, Water A was selected. (See Table 2 for composition).

The evaporator design is based on removing approximately 90 percent of the water contained in the feed. The crystallizer removes most of the remaining water and delivers a slurry of crystals to the waste pond.

4.3.2 Test Experience

Vertical tube evaporators have been tested for several years in the desalination of seawater. A 1 mgd plant was operated in the downflow mode for approximately two years at Freeport, Texas, using the falling film technique.(Ref. 2 and 3) In addition, vertical tube evaporation has been widely used in crystallization processes in the so-called Oslo type crystallizers in which, however, the flow pattern more closely approximates full-tube flow rather than film flow.

No industrial applications of the VTFE exist at present. A 5,000 gallon per day pilot plant evaporator-crystallizer has been subjected to extended tests at the Sea Water Conversion Laboratory (SWCL) of the University of California at Richmond, California.(Ref. 4)

4.3.3 Process Description

As in the case of the vapor compression evaporator, the VTFE is basically a vertical tube evaporator. Two modes of operation have been tested. In one mode, the feed (cooling tower blowdown in several tests performed at the SWCL) is pumped to the head of the evaporator, is distributed across the tube sheet, and enters the tubes via specially designed nozzles, permitting it to fall as a film on the inner walls of heat exchanger tubes. In the other mode, the feed enters the sump at the bottom of the evaporator, passes through a perforated distributor plate, and rise a short distance up the heat exchanger tubes. In both modes, steam condensing on the outside of the tubes causes evaporation of the liquid within the tubes. In the second mode, referred to as upflow operation, steam bubbles generated within the liquid in the bottom portion of each tube tend to rise, carrying with them some liquid. This liquid is ejected from the upper ends of the tubes, resembling somewhat the action of a coffee percolator.

The upflow mode of operation is advantageous because no recirculating pump is required, either in the single effect design under study here or for the transfer

of brine to each succeeding effect in a multieffect plant. In addition to the saving in plant investment, energy cost is also reduced. However, at low operating temperatures such as those encountered in systems using waste heat, the hydraulic driving force may not be adequate for upflow operation even after surfactant is added. Therefore, at the low temperatures anticipated in the field tests of this study, the downflow mode will be used.

For either mode of operation, the addition of a few ppm of surfactant results in improved performance. In downflow, the additive assists in distributing the liquid as a uniform thin film on the inner tube wall. In upflow, a suitable surfactant stabilizes the foam produced in the tubes as a result of steam generation, improving the operational stability and heat transfer coefficient and decreasing hydrostatic pressure loss. A number of foaming surfactants have been tested successfully at SWCL; an alkyl benzene sulfonate appears to be preferable at present.

4.3.4 Plant Design

The flow diagram for the vertical tube foaming evaporation system is shown in Figure 12. Six modules are provided, each consisting of an in-line static mixer, vacuum deaerator, evaporator, recirculating pump, and condenser. Raw cooling tower blowdown is fed to each module at 250 gpm. The pH of each stream is adjusted from 7.3 to within the range of 5.5 to 6.0 by the addition of 66^{o}Be sulfuric acid. After acid injection, the feed stream flows through a 3 inch diameter, 1 foot long, 316L stainless steel in-line static mixer. The acid addition rate is regulated by the pH reading in the static mixer effluent.

The stream next flows to a 4 foot diameter vacuum deaerator where oxygen, nitrogen, and carbon dioxide are removed and vented to the atmosphere. The deaerator shell is fabricated from 316L stainless steel. After leaving the deaerator, the stream flows into the suction line of the evaporator recirculation pump, which recirculates brine at 6,000 gpm to the tube bundle at the top of the evaporator. Here, a perforated plate distributes the flow across the tube bundle, which consists of 2,600 titanium tubes, each of which is 26 feet long with an outside

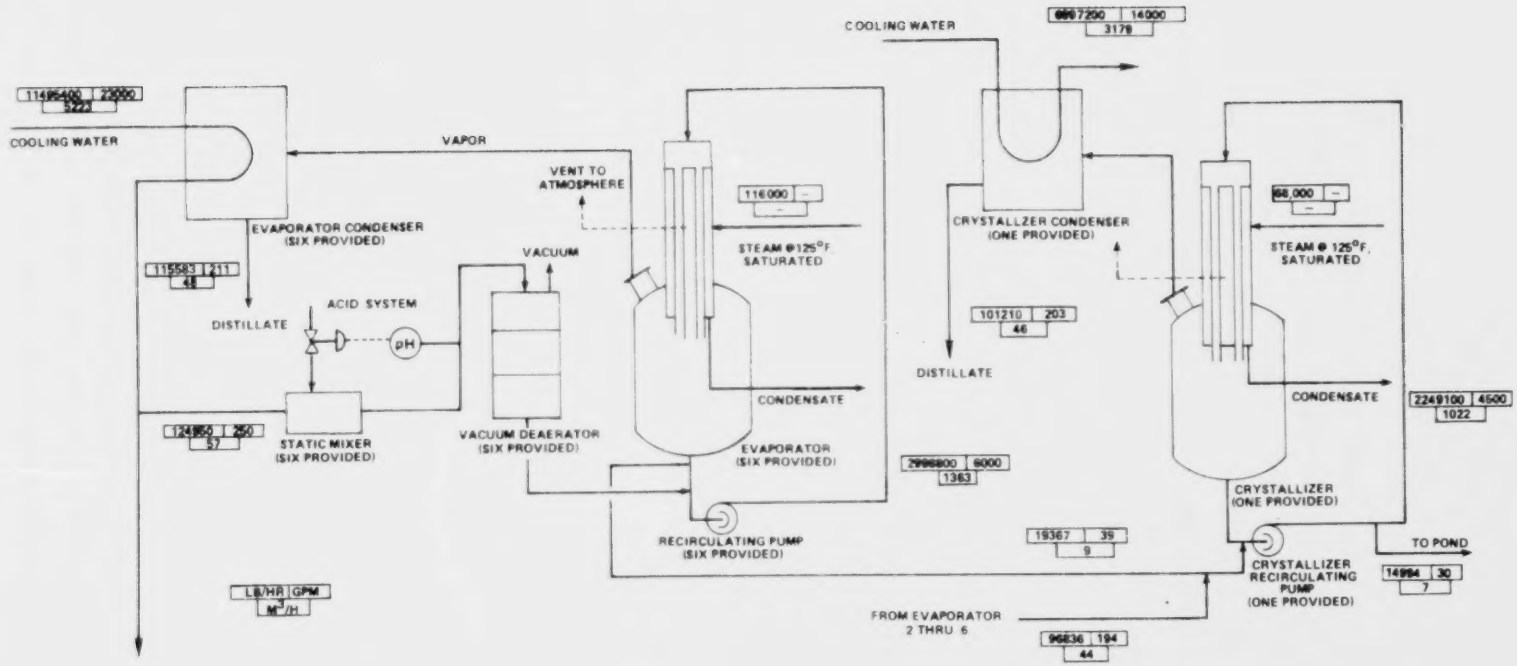


Figure 12. Vertical tube foaming evaporation option — downflow mode.

diameter of 1.5 inches. The tube bundle is enclosed in a 106 inch diameter 316L stainless steel shell. The brine flows down the inner walls of the tubes as a thin film and is heated by steam, which is supplied to the tube bundle at 116,000 pounds per hour. The temperature of the steam is 125°F. A mixture of vapor and liquid descends into a 316L stainless steel evaporator, which measures 36 feet in diameter by 12 feet. Operating conditions in the evaporator are 105°F and 55 mm absolute pressure. Noncondensibles are removed by a steam ejector supplied with 700 pounds per hour of steam at 90 psig. The liquid in the evaporator returns to the recirculation pump and the vapor passes through a mesh demister to a shell-and-tube condenser.

The condenser has 5,344 titanium tubes, each 18 feet long with an outside diameter of 1 inch. The condenser shell measures 100 inches in diameter and is fabricated from 316L stainless steel. Cooling water with a temperature of 80°F flows through the condenser at 23,000 gpm. The cooling water leaves the condenser at 90°F, and a 250 gpm stream is diverted to feed the evaporator. Condensate leaves the deaerator at 211 gpm.

As the brine is recirculated to the top of the evaporator, a 39 gpm stream is diverted from each evaporator to make up the feed stream to the crystallizer. The streams are combined and enter the suction line of the crystallizer recirculating pump, which delivers brine to the top of the crystallizer at 4,500 gpm. Here, a perforated plate distributes the flow across the tube bundle, which consists of 1,950 titanium tubes, each 26 feet long with a 1.5 inch outside diameter. The tube bundle is enclosed in a 92 inch diameter 316L stainless steel shell. As with the evaporator, 125°F steam is fed to the steam chest of the crystallizer to heat the brine as it flows down the inner walls of the tubes. The steam is supplied at 68,000 pounds per hour. A mixture of liquid and vapor falls into the crystallizer body, which is 27 feet in diameter by 12 feet. The crystallizer operates at 109°F and a pressure of 55 mm absolute. Noncondensibles are removed by a steam ejector. The steam requirement for the ejector is 700 pounds per hour 90 psig steam. The liquid in the crystallizer returns to the recirculating pump, and the vapor passes through a mesh demister to a shell-and-tube condenser.

The crystallizer-condenser has 3,150 titanium tubes, each 18 feet long with an outside diameter of 1 inch. The condenser shell measures 61 inches in diameter and is fabricated from 316L stainless steel. The cooling water requirement for the condenser is 14,000 gpm of 80°F water. Condensate is produced at 203 gpm.

The system design was based on a heat transfer coefficient of 250 Btu/hour·sq ft·1°F. A 30 gpm stream is diverted from the recirculating brine at the crystallizer and discharged to 50 acre solar evaporation pond. The pond size is based on a net evaporation rate of 4.0 feet per year.(Ref. 1) The water depth in the pond is 10 feet. A half acre emergency pond is provided to store three days' flow to one of the six evaporators.

The layout of the vertical tube foaming evaporation plant is shown in Figure 13.

4.3.5 Process Economics

No vertical tube foaming evaporation system is in use at present, nor are any field data available to permit a realistic cost estimate of a commercial size system at the time this report is prepared. Consequently, the economics presented here is based on a conventional vertical tube evaporator plant designed to operate with 125°F steam, as agreed by the Environmental Protection Agency. This low temperature imposes a cost penalty resulting from the large steam chest and line required by the high specific volume of steam. The plant cost for the VTE, which serves as the basis for this estimate, represents the mean of two quotations from established suppliers.

To attain a 30 year plant life without the need for major repairs or replacement, the quotations were based on the use of smooth titanium for the evaporator tubes and 316L stainless steel for all portions of the equipment in contact with brine. This is admittedly conservative, since wetted parts of carbon steel could be used under carefully controlled operating conditions. However, numerous cases of failure of carbon steel components in evaporators indicate that such careful control is seldom exercised in practice. Therefore, the conservative approach used in this study duplicates the material selection employed in current vapor compression plants and used as the basis for Subsection 4.2.4 of this report.

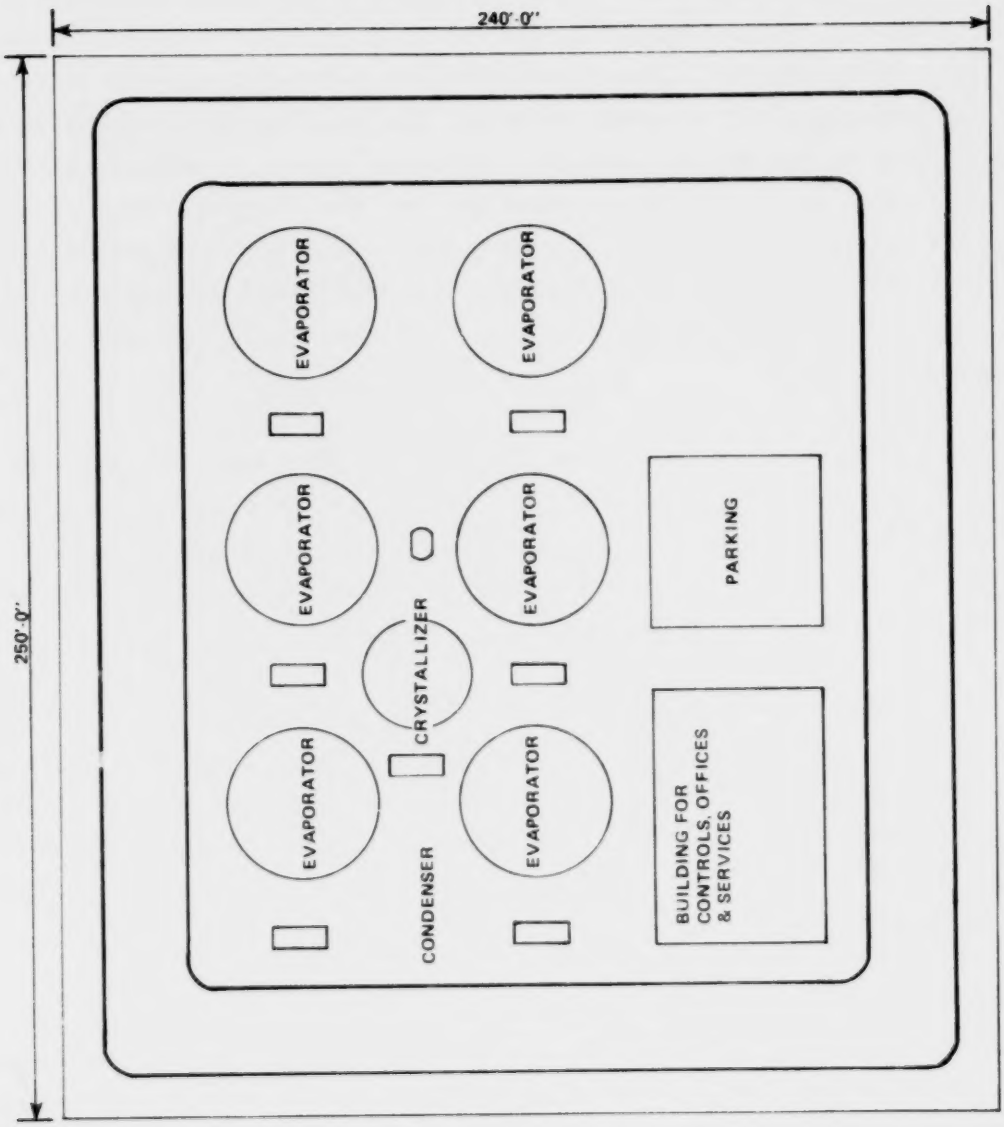


Figure 13. Plot plan for vertical tube foaming evaporation option.

The total installed cost of the VTFE plant, including ponds for the storage of blowdown, was estimated at $15,900,000.

The overall heat transfer coefficients that serve as the basis for evaporators quoted to Bechtel ranged from 225 to 250 Btu/hr·ft^2·1^{o}F. A value of 250 was assumed for this analysis. This is substantially below the value of 500 postulated for commercial VTFE evaporators by the Sea Water Conversion Laboratory of the University of California. However, their coefficient is based on pilot plant experience at the Laboratory where the feed was reconstituted. If a coefficient of 500 can be substantiated by prolonged operation under actual field conditions without frequent tube cleaning, then the plant cost presented in this study would decrease by approximately 8 percent and the treatment costs by 6 percent.

When heated by turbine exhaust steam, the VTFE in effect takes the place of a portion of the power plant condenser. In a newly designed plant, as opposed to a retrofit, the VTFE appears to reduce the size and cost of the condenser. In practice, however, the full condenser would be installed to handle the exhaust steam in the event of a VTFE shutdown.

A credit has been taken indirectly for the condensing function of the VTFE: in this study, the plant has not been charged for the energy required to deliver the cooling tower blowdown to the evaporator or for returning the distillate to the cooling tower. Instead, it was assumed that it would merely function as a bypass for some of the coolant normally pumped from the cooling tower to the power condenser.

Because of the assumption of a no-cost heat source, the energy consumption charged to the process is quite low, consisting chiefly of pumping power requirements. Consequently, the dependence of treatment cost on the price of power, as shown in Figure 14, is very small. Also, as in the case of the VC plant discussed in Subsection 4.2.5, only 2 percent of the water content of the cooling tower blowdown is assumed to require ponding. Thus, the slope of the "treatment cost vs. pond cost" curves in Figure 15 is very low.

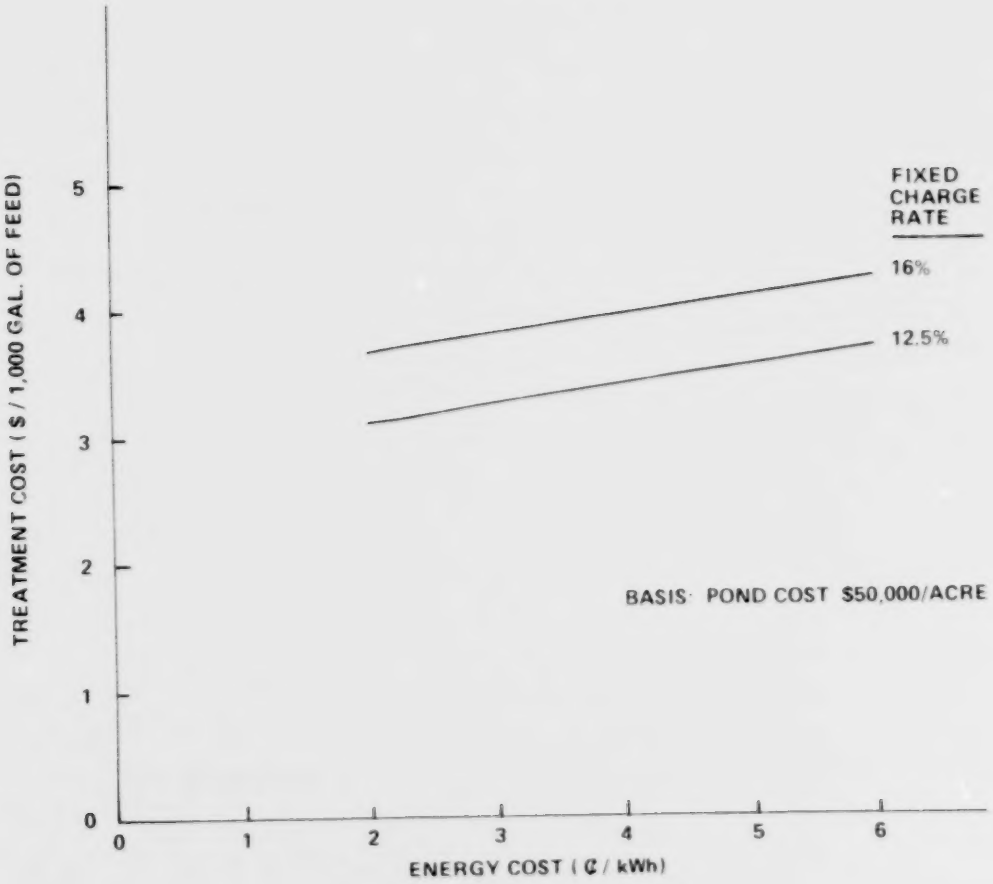


Figure 14. Effect of energy cost on VTFE treatment costs.

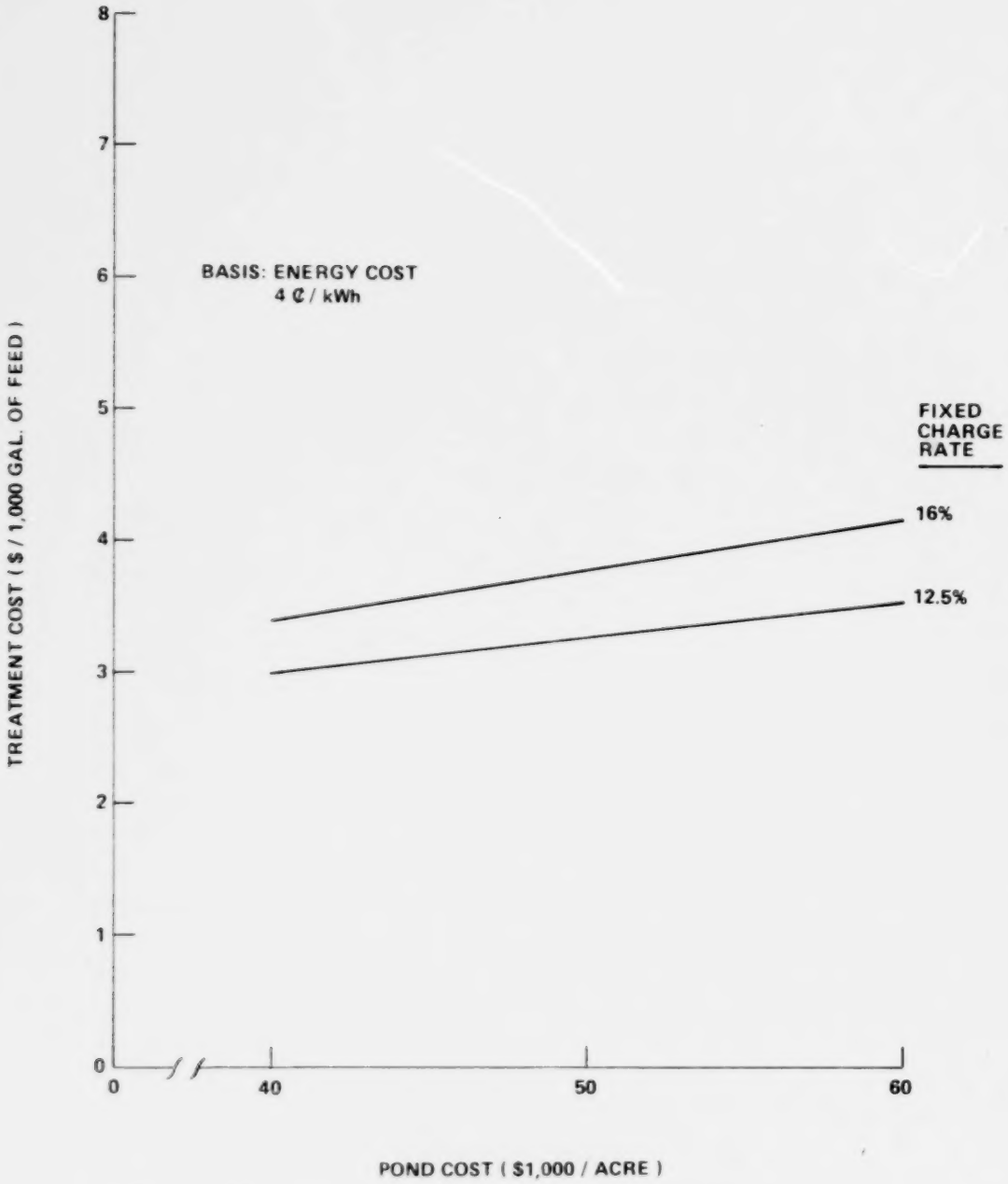


Figure 15. Effect of pond cost on VTFE treatment costs.

If it is subsequently decided, for the purpose of upflow operation or for other reasons, to operate the VTFE at higher temperatures, the heat will no longer be "free." Steam at higher temperatures can be provided by extraction from the turbine at some point above the normal exhaust pressure. A value can be assigned to the extraction steam by the use of curves in Reference 5, updated to reflect current costs, and the costs in this report could be adjusted accordingly.

No charge was included to cover the cost of the surfactant which, at the low concentrations tested to date, is comparatively minor. It has been assumed that the surfactant accompanies the blowdown stream to the solar evaporation pond, where its presence is not objectionable. If a particular blowdown stream contains salable salts, it may be necessary to free it of surfactant. In that case, the removal cost would have to be balanced against the value of the recovered salts.

The VTFE costs are compared with those of the other two processes in Subsection 4.5.

4.4 OVERALL PLANT WATER BALANCE

In the concentration of cooling tower blowdown, all of the processes analyzed in this report produce a supply of purified water suitable for many uses in the power plant. Among the potential uses are:[Ref. 6]

- Fossil-fired power plants need water for the scrubbers of the flue gas desulfurization system. This represents a comparatively minor requirement, and the water used here can be high in dissolved solids, provided that it is not excessively scaling in nature.

- Coal burning plants consume rather small amounts of water, again not necessarily of high quality, for ash sluicing.

- Boiler makeup and water for flushing pump glands represent, at most, 5 percent of the power plant water demand. Since water of very high purity is required here, two-bed ion exchange followed by mixed bed polishing is commonly employed. For very pure feeds, such as the distillate from the VCE or VTFE process, only the polishing step is required.

- The major water requirement is for the cooling system. In general, most of the feed to the cooling loop is for evaporative cooling. The residual water constitutes the cooling tower blowdown.

For simplicity of this study, it has been assumed that all the water recovered by cooling tower blowdown concentration is returned to the cooling tower. This procedure has a threefold advantage:

- The quantity of makeup water to the cooling system is reduced, thereby decreasing the cost of raw water treatment. Although the evaluation of raw water treatment was considered to be outside the scope of the present study, it should be considered in a future, more detailed analysis

- Since a fraction of the feed to the cooling tower loop is pure distilled water, less dissolved salt is introduced and, hence, the flow rate of cooling tower blowdown is reduced. Consequently, a smaller and less costly plant is required for the concentration of the blowdown

- The rate of discharge of brine from the concentrator is iecreased. This results in a lower ponding requirement.

The comparison of process conditions for operation with and without recycle of purified water is presented in the figures for RO and distillation processes. Figures 16 and 17 use Water A and Water B, respectively, in a reverse osmosis system. Figure 18 shows the effect of water recycle in a distillation system using Water A. (See Tables 2 and 3 for the composition of feed waters A and B.)

The calculations are based on the following assumptions:

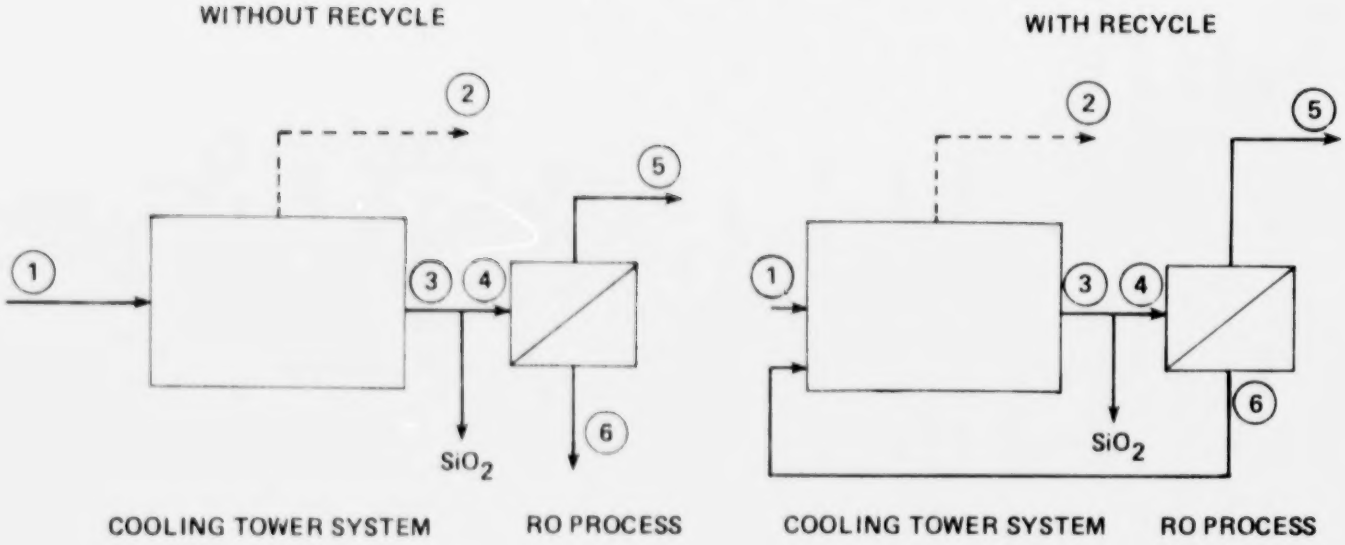
- The cooling tower blowdown rate in all cases represents one-fourth of its makeup flow and, consequently, the concentration of dissolved solids ("cycles" of concentration) is four times that in the makeup
- Either distillation process, VCE or VTFE, recovers 98 percent of the water of the cooling tower blowdown
- The RO process recovers 75 percent of the water of the cooling tower blowdown
- The salinity of the RO product is not altered by the water recycle scheme
- For the reference power plant, the cooling tower evaporation rate is not influenced by recycle of product water from the concentrator.

From the figures, it is apparent that the recycle of product water from the concentrator to the cooling tower when RO is used has the following advantages:

- A 20 percent reduction in makeup water required for the power plant cooling system (Column 1)
- A 16 percent reduction in the size of the plant for pretreating and concentrating the cooling tower blowdown (Columns 3 and 4)
- A 16 percent reduction in the flow of reject brine to the pond (Column 5).

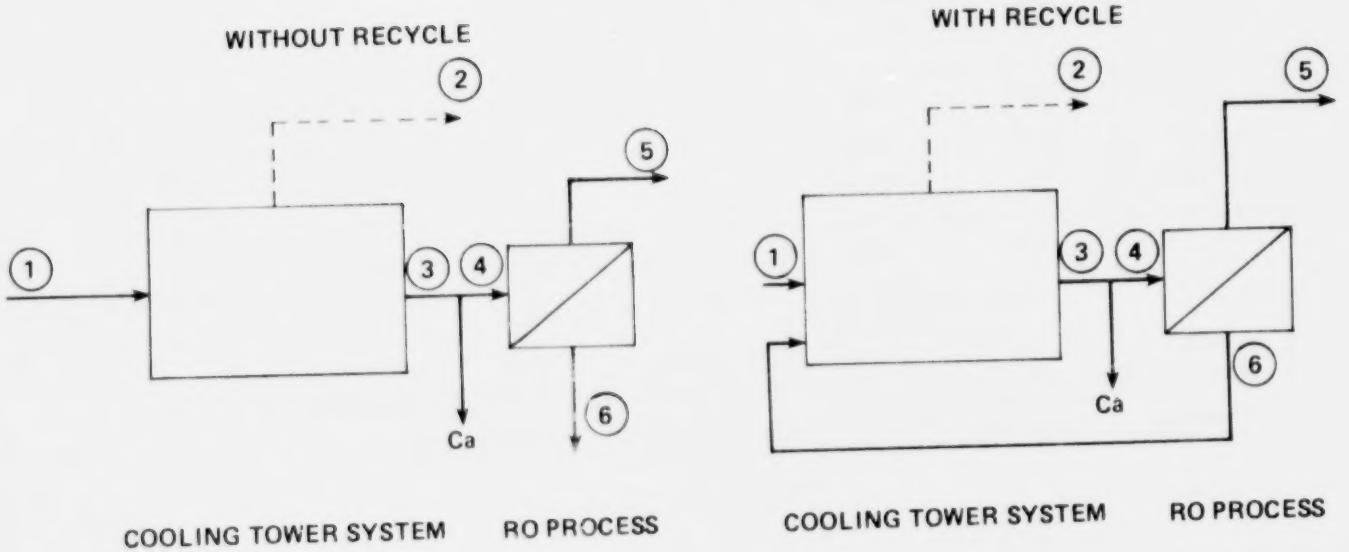
For the distillation processes, the advantages are:

- A 25 percent reduction in the power plant coolant makeup (Column 1)
- A 25 percent reduction in the size of the evaporation plant (Column 3)



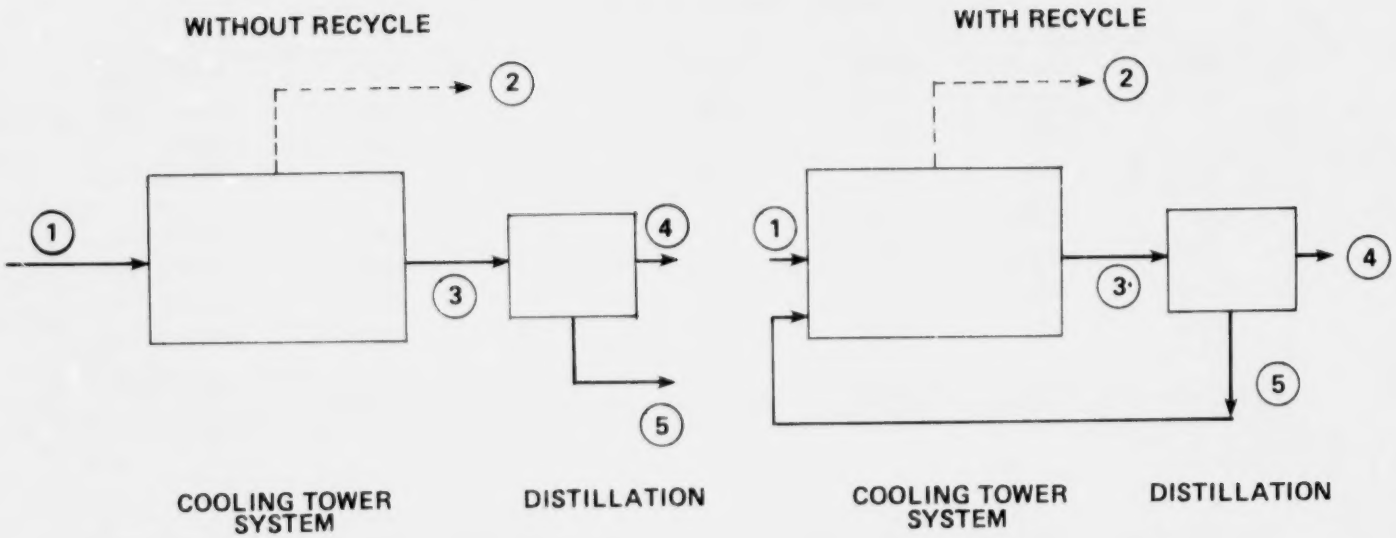
PROCESS	STREAM / PARAMETER	MAKEUP 1	EVAPORATION 2	BLOWDOWN 3	TREATED BLOWDOWN 4	REJECT TO POND 5	PRODUCT 6
WITHOUT RECYCLE	FLOW (gpm)	8000	6000	2000	2000	500	1500
	TDS (ppm)	1880	0	7522	7619	28,855	475
	SiO_2 (ppm)	37.5	0	150	37.5	125	8.5
WITH RECYCLE	FLOW (gpm)	6420	6000	1700	1700	425	1275
	TDS (ppm)	1880	0	7500	7550	28,855	475
	SiO_2 (ppm)	37.5	0	150	37.5	125	8.5

Figure 16. Comparison of flows with and without recycle of purified water using Water A and reverse osmosis.



PROCESS	STREAM / PARAMETER	MAKEUP 1	EVAPORATION 2	BLOWDOWN 3	TREATED BLOWDOWN 4	REJECT TO POND 5	PRODUCT 6
WITHOUT RECYCLE	FLOW (gpm)	8000	6000	2000	2000	500	1500
	Ca (ppm)	133	0	533	50	200	0
	SO_4 (ppm)	955	0	3820	3847	14,638	214
	TDS (ppm)	1391	0	5562	5650	21,358	384
WITH RECYCLE	FLOW (gpm)	6410	6000	1700	1700	425	1275
	Ca (ppm)	133	0	533	50	200	0
	SO_4 (ppm)	955	0	3890	3920	15,000	214
	TDS (ppm)	1391	0	5730	5820	22,100	384

Figure 17. Comparison of flows with and without recycle of purified water using Water B and reverse osmosis.



PROCESS	STREAM / PARAMETER	MAKEUP 1	EVAPORATION 2	BLOWDOWN 3	BRINE TO POND 4	PRODUCT 5
WITHOUT RECYCLE	FLOW (gpm)	8000	6000	2000	40	1960
	TDS (ppm)	1880	0	7522	SAT	0
	SiO_2 (ppm)	37.5	0	150	SAT	0
WITH RECYCLE	FLOW (gpm)	6031	6000	1500	30	1470
	TDS (ppm)	1880	0	7520	SAT	0
	SiO_2 (ppm)	37.5	0	150	SAT	0

Figure 18. Comparison of flows with and without recycle of purified water using Water A and distillation.

- A 23 percent reduction in the flow of distillation plant blowdown to the pond (Column 4).

As a result of the above comparisons, the recycle of product water from the blowdown concentrator back to the cooling tower has been incorporated in each of the three process flow sheets in Subsections 4.1, 4.2, and 4.3.

4.5 COMPARISON OF PROCESSES

Before comparing processes having characteristics as diverse as the three processes analyzed in this report, it was necessary to establish a common basis for the analysis. The problems stem from the fact that two of the processes use mechanical energy, and the third uses heat which, under the conditions postulated here, is considered to have zero cost. In addition, the membrane process produces a purified water stream substantially inferior in total dissolved solids content to the other two.

To compensate for the differences in product water quality, the water was assumed to recirculate to the cooling tower in every case. Thus, the process that delivers the purer product is automatically credited by virtue of the resulting decrease in makeup to the cooling tower system and in the size of plant required to treat the cooling tower blowdown. These relationships are apparent in Figures 16, 17, and 18 and in Table 8. The ponding requirements of the processes studied here were taken into consideration by calculating the pond area required for a 30 year plant life and considering this cost as part of the investment in the cooling tower blowdown treatment plant.

Table 9 summarizes the costs and energy consumption of the three processes. The values listed there are based on an assumed pond cost of $50,000 per acre, an energy cost of 4¢ per kWh, and a 12.5 percent fixed charge rate. From the standpoint of plant investment, reverse osmosis is definitely superior to the two distillation processes. In total operating cost, RO is only slightly inferior to the VTFE.

If it becomes necessary to raise the operating temperature of the VTFE, the favorable energy consumption of RO would become even more pronounced by comparison, and its operating cost could conceivably be lower than that of the VTFE.

These and other attributes of the three processes are listed in Table 10. From this comparison, it appears that no process is definitely superior to the other two in all respects. The picture promises to change with time and as a result

of further work. For example, successful field testing of the VTFE will provide confidence in the process, for which there is currently no basis. It could also result in more favorable heat transfer rates than were conservatively assumed in this report.

It was concluded from the comparison in Table 10, that analysis should continue in order to examine the merits of a combination process. This would involve concentration of cooling tower blowdown by a membrane process, which will deliver its reject stream to a distillation plant to minimize the volume of liquid waste requiring ponding.

TABLE 8. EFFECT OF RECOVERY AND PRODUCT PURITY ON SYSTEM DESIGN

	Reverse Osmosis	Distillation
Requirement for makeup water to power plant cooling system (gpm)	6,420	6,031
Cooling tower blowdown requiring treatment (gpm)	1,700	1,500
Blowdown flow from treatment system (gpm)	425	30
Ponding area required for 30-year plant life (acres)	160	50

Basis: 700 MWe power plant.
Ponds 10 feet deep.

TABLE 9. COMPARISON OF COSTS AND ENERGY CONSUMPTION

	Reverse Osmosis	Vertical Tube Foaming Evaporation	Vapor Compression
Plant investment, 1st quarter '79 (million dollars)			
Treating plant	3.7	12.5	11.8
Ponds	10.2	3.4	3.4
Total	13.9	15.9	15.2
Annual cost (million dollars per year)(1)			
Capital cost	1.7	2.0	1.9
O&M cost			
Supplies	0.7	0.1	0.1
Electrical energy	0.2	0.3	2.6
Labor	0.3	0.2	0.2
Total O&M cost	1.2	0.6	2.9
Total annual cost	2.9	2.6	4.8
Energy consumption(2)			
Electricity (million kWh/year)	5.3	6.7 (not free)	63.9
Steam (10^{12} Btu/year)	—	7.0 (no cost)	—

(1) Basis: 30-year plant life, ponds at $50,000/acre, electricity at 4¢/kWh, 12.5 percent fixed charge rate. See above for a discussion of effects of variations in these assumptions.
(2) Energy cost is included in O&M cost above.

Note: The vertical tube foaming evaporation (VTFE) costs are estimates based on VTE experience. They may change substantially as VTFE operating data become available.

TABLE 10. OVERALL COMPARISON OF PROCESSES

	Reverse Osmosis	Vertical Tube Foaming Evaporation	Vapor Compression Evaporation
Ponding requirements[1]	Large	Moderate	Moderate
Long term-related operating experience on cooling tower blowdown	Minimal	None	Considerable and successful
Modular construction of sub-units of each train	Standard procedure	Not practical	Not practical
Sensitivity to upsets in pretreatment	Very sensitive	Minor	High
Product quality	Fair (475 ppm)	High (10 ppm)	High (10 ppm)

(1) Ponding cost is included in the plant investment and the calculation of annual cost in Table 9.

SECTION 5

TEST PLAN FOR VTFE SHAKEDOWN TESTS

The following test plan for the shakedown of the VTFE at the Sea Water Conversion Laboratory (SWCL) was developed with the cooperation of Dr. Hugo H. Sephton, principal investigator at the SWCL. See Figure 19 for a photograph of the pilot plant. The tests are directed toward assuring the structural integrity and the safe and smooth operation of the pilot plant equipment. Additional improvements in the equipment and refinements in operation and controls may result from the field test program which is to follow this shakedown phase.

Shakedown testing will be performed and improvements in process control implemented for each of the facility units, i.e., boiler, VTE evaporator, VTE crystallizer, and condenser. The vacuum tightness of all welds will be checked. Process control elements will include: steam temperature control; brine distribution; adequate bundle venting, couplings, valves, etc.; coolant flow and temperature control for controlling the ΔT applied; boiler controls and steam desuperheating; condensate and distillate removal and flow measurement; temperature measurement and recording; and startup and shutdown procedures. Calibration will be performed on all devices for measuring process parameters.

To protect the pilot plant, temperature sensors will be provided. The sensors will actuate high- and low-temperature alarms and will shut the plant down in the event of temperature excursions. Level controllers will control either the feed or the blowdown rate of the evaporator in reponse to the level in the evaporator sump. The variable not regulated by the level controller will be regulated by a salinity monitor. A level controller will regulate the rate of discharge from the crystallizer.

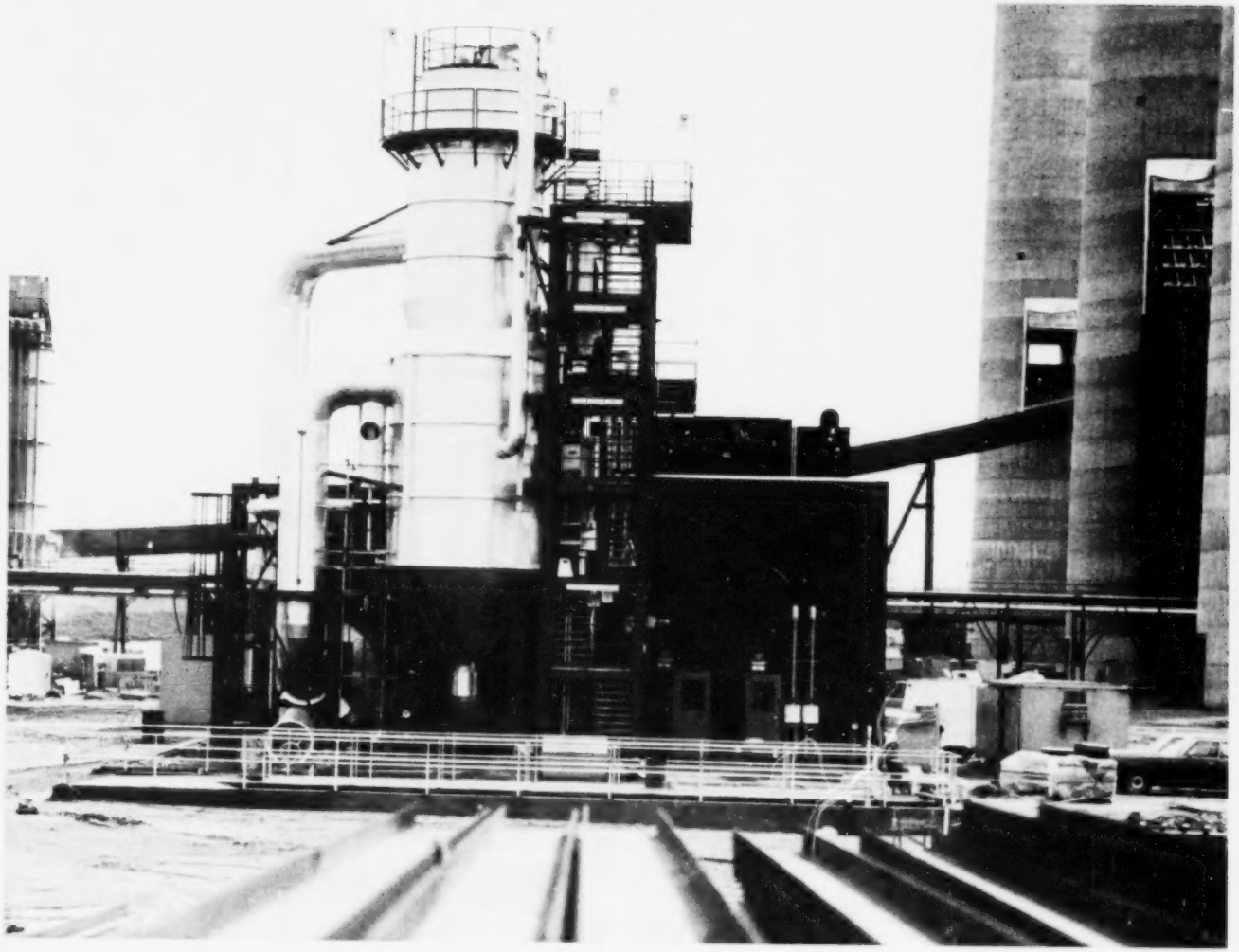


Figure 19. Photograph of VTFE pilot plant.

Three types of feed waters will be tested during the shakedown period:

- Pure water (For convenience, tap water will be used.)
- Seawater diluted with tap water to approximate a cooling tower blowdown composition
- Diluted seawater containing the preferred surfactant.

The surfactant tests have been relegated to the end of the shakedown tests, since previous experience shows that it is very difficult to remove surfactants from the system once they have been injected.

With each of these feeds, the feed liquor will be reconstituted by combining the distillate and blowdown in order to minimize the cost of the shakedown tests. It is realized, however, that reconstituting the feed does not present a realistic picture of sludge and scale formation or of corrosion phenomena. Consequently, it is anticipated that once field testing is initiated at a power plant site, an actual cooling tower blowdown will be used as feed and the concentrate will be discarded or ponded.

A matrix of tests will be instituted, incorporating the following variables:

1. Operating temperature

   a. Low temperature (about 125°F), representing waste heat from a condensing type of steam turbine

   b. Medium temperature (about 150°F), representing heat from steam discharged from a back-pressure turbine

   c. High temperature (about 220°F), as would be encountered in the use of prime steam after pressure reduction and desuperheating.

2. Flow regime

   a. Upflow, using the flashing of the brine to drive the two-phase fluid upward and out of each tube (Note that conditions 1a and 1b above many not provide sufficient hydraulic driving force to permit upflow operation. Measurements of pressure drops across the orifice plate and up the tubes will define the limits of upflow operation possible without pumps.)

b. Downflow, using a recirculation pump to raise the brine to the plenum that supplies the evaporator tubes.

3. Surfactant addition

a. Without surfactant, to establish a base line for comparison

b. With surfactant, using the type of surfactant and concentration demonstrated as optimum in the preliminary studies performed at the Sea Water Conversion Laboratory.

Under 2b (downflow), the effect of flow rate of the brine should also be investigated. This, in turn, controls the thickness of the fluid film on the interior surfaces of the tubes. Conversely, in upflow the liquid level in the evaporator tubes controls the mass flow rate through the tubes and, consequently, will be varied to the extent possible.

The measurements required, as a minimum, are:

- Evaporator
  - Steam supply
    - Pressure
    - Temperature
    - Flow of condensate
  - Brine evaporation
    - Feed: Temperature
      Concentration of dissolved solids
      Flow rate
      pH
    - Vapor: Temperature
      Pressure
      Venting rate
    - Concentrate: Temperature
      Concentration of dissolved solids (salinity)
      Concentration and composition of suspended solids
      pH

- Condenser
  - Condensate
    - Temperature
    - Flow rate
    - Purity (in tests with saline feeds only)*
  - Coolant
    - Temperature entering
    - Temperature leaving
    - Flow rate
- Crystallizer
  - Steam supply
    - Pressure
    - Temperature
    - Flow of condensate
  - Brine evaporation
    - Vapor: Temperature
      Pressure
      Venting rate
    - Concentrate: Temperature
      Concentration (salinity)
      Flow rate

In addition, it is essential to determine the uniformity of brine distribution to the evaporator tubes and the pattern of steam disengagement in the evaporator.

Quantitative data are highly desirable if attainable. At the very least, records will be kept of periodic visual observations, preferably supported by photographs.

*If possible, it is desirable to obtain separate data on flow rate and purity of vapors from both the evaporator and the crystallizer.

Uniform brine distribution will not only improve evaporator efficiency, but will avoid the possibility of baked deposits in the tubes. As for steam disengagement, a poor velocity pattern will cause carryover of a large quantity of brine along with the vapor.

As a check, a material balance will be prepared, comparing the feed entering the evaporator and the concentrate leaving the evaporator. A similar material balance will be calculated around the crystallizer. Such material balances serve as cross-checks on flow measurements and analytical results. They have the additional function of providing an early indication of buildup of deposits in the lines and equipment.

Finally, whenever the equipment is opened, sludge and scale deposits will be removed, examined visually, weighed, and analyzed.

The data enumerated above will:

- Permit the calculation of mean effective overall heat transfer coefficients
- Permit separate calculation of economy ratio of the evaporator and the crystallizer
- Alert the operator to the scaling or fouling of the condenser tubes
- Give an indication of the malfunctioning of the plant components
- Permit the calculation of the thickness of the evaporating film of liquid on the interior tube walls
- Serve as a guide in the "fine tuning" of the pilot plant

The longevity and the economics of an evaporator-crystallizer are strongly influenced by corrosion. Consequently, the following tests will be performed before startup and after completion of the shakedown tests:

- Measure the wall thickness of evaporator and crystallizer tubes from several representative locations (near the outside of the

bundle, near the center, and at some intermediate location). Such measurements will be made on the portions of the tubes that protrude from the upper tubesheet. Measurements will be performed before the start and at the end of the shakedown testing and will be recorded together with visual observations of the tube surfaces.

- Install, in the vessels and lines, coupons representing the materials of construction of the plant and also a few potentially desirable alternatives. Two types of coupons are required. One group will consist of unstressed coupons, weighed and measured before installation and again at the end of the shakedown tests. The second group will consist of stressed U-bend coupons of the same materials.

- Measure the thickness of the pump impeller at several specific loations. Whenever the plant is opened, inspect for rust and pitting and take photos.

- With inside calipers, check changes in the inside diameter of piping in selected locations.

- Determine corrosion/erosion of elbows by observing them at the start and end of shakedown testing. Several nondestructive testing methods are available, among them radiography and ultrasonic testing. Efforts will be made to locate sources of rental equipment or testing services to determine the most economical procedure.

The above program is directed toward the shakedown phase of the overall test program. Many of the tests outlined in this section, however, are suitable for continued monitoring during field testing and will be repeated at the termination of work at each power plant site.

The shakedown program was initiated during the last two weeks of activity under EPA Contract No. 68-02-2616 with Bechtel. During the shakedown, the pilot evaporator was tested in the upflow mode, the crystallizer in the downflow mode. (Please refer to Subsection 4.3.) The feed consisted of fresh water. The flow up the evaporator tubes was accomplished with the aid of a pump, in contrast to the hydrothermal driving force planned for the field tests.

In tests witnessed by Bechtel, the evaporator feed consisted of a mixture of coolant water and distillate. The coolant water was preheated by passing through the vent condenser of the evaporator to a temperature approaching that of the recirculating brine, which was at 111.3°F. The preheated coolant water and distillate were mixed with the recirculating brine and then delivered to the evaporator at the rate of 600 gpm. No acid or surfactant was added to the feed.

An oil fired boiler delivered steam at 25 psig. The steam pressure was reduced by a temperature-controlled throttling valve and desuperheated by the injection of a spray of preheated condensate. Excess moisture was stripped from the steam in a cyclone separator designed at the SWCL. The temperature of the desuperheated steam was 132.3°F, and the temperature of the saturated steam in the steam chest was 131.8°F.

The vapor generated by evaporation of the feed in the evaporator (and of the concentrate in the crystallizer when heated) was condensed in the two-pass condenser, where heat was rejected to cooling water flowing through the tubes at the rate of 1,401,120 pounds per hour. The coolant temperature was maintained at 107.5°F by varying the ratio of fresh coolant to water recycled from a sump. Distillate was produced at a rate of 10 gpm.

By visual observation through sight glasses in the vapor dome, it was concluded that the recirculating brine was distributed fairly uniformly among the evaporator tubes. Under the operating conditions tested, the evolved vapor did not appear to produce excessive scatter of the droplets of liquid emerging from the evaporator tubes.

Operation of the plant was quiet and steady. The liquid levels and salinity were maintained reasonably constant by the control valves in the respective feed and blowdown lines. The pH and salinity were recorded automatically on strip charts. Plant startup was relatively simple; equilibrium conditions were attained in approximately 90 minutes from firing of the boiler. Shutdown was uneventful.

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8) Langelier, W. F., "Chemical Equilibria in Water Treatment," Jour. AWWA, 38, No. 2, 169-178 (February 1946).

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Appendix A

BASIC WATER CHEMISTRY

The composition of water delivered to the cooling system of a power plant depends on plant location and, in addition, is subject to seasonal variations. Of all the constituents of raw water, the most important from the standpoint of both the cooling tower and the blowdown treatment system are the scale formers. Three types of scale merit special consideration: calcium carbonate, calcium sulfate, and silica.

CALCIUM CARBONATE SCALING

The bicarbonate ion, which is present in many water supplies, is in equilibrium with dissolved carbon dioxide:

$$2HCO_3^- \rightleftharpoons CO_3^{--} + H_2O + CO_{2(DIS)} \qquad (1)$$

where $CO_{2(DIS)}$ represents molecular carbon dioxide dissolved in the water. The dissolved carbon dioxide, in turn, is in equilibrium with CO_2 in the atmosphere:

$$CO_{2(DIS)} \rightleftharpoons CO_{2(GAS)} \qquad (2)$$

When water containing bicarbonate ion contacts air in the cooling tower, the dissolved CO_2 is reduced to a value determined by the Henry's Law coefficient and this, consequently, drives reaction (1) to the right. The resulting increase in carbonate ion, together with the increased concentration caused by evaporation, results in the deposition of calcium carbonate scale in the cooling tower and associated piping:

$$Ca^{++} + CO_3^{--} \rightleftharpoons CaCO_3 \qquad (3)$$

After depositing the scale, the water is still saturated in $CaCO_3$, which has a negative temperature coefficient of solubility. Thus, when the water is heated

by passage through the power plant condenser, further deposition of $CaCO_3$ scale occurs, but now in the condenser tubes.

One solution to this problem, used in certain types of distillation processes, is the destruction of the bicarbonate ion by acid addition. In this process, sulfuric acid is preferred for acidification because of its low cost:

$$2HCO_3^- + H_2SO_4 \rightleftharpoons SO_4^{--} + 2H_2O + 2CO_{2(DIS)} \quad (4)$$

The dissolved CO_2 is stripped from the acidulated water, in accordance with reaction (2), by a countercurrent stream of air in the decarbonator.

A commonly used criterion for the stability of water against calcium carbonate precipitation was developed by Langelier.(Ref. 7, 8, 9) The Langelier Index is defined by the equation:

$$L.I. = pH_m - pH_s, \text{ where} \quad (5)$$

L.I. = Langelier Index

pH_m = the measured pH of the water in question

pH_s = the "saturation" pH.

The "saturation" pH was shown by Langelier to be:

$$pH_s = (pK_2' - pK_s') + pCa + pAlk, \text{ where} \quad (6)$$

$(pK_2'-pK_s')$ = a correction depending primarily on the ion strength of the solution

Ca = the molar concentration of Ca ions present.

Alk = the alkalinity, expressed as moles per liter of $CaCO_3$ corresponding to the CO_3^{--}, HCO_3^-, and OH^- in the solution

p = indicates the logarithm of the reciprocal of the quantity indicated. For example, $pCa = \log (1/ [Ca^{++}])$.

Both pCa and pAlk can be read directly from Figure A-1. The value of $(pK_2'-pK_s')$ can be found in Table A-1, which takes the effect of temperature into consideration.

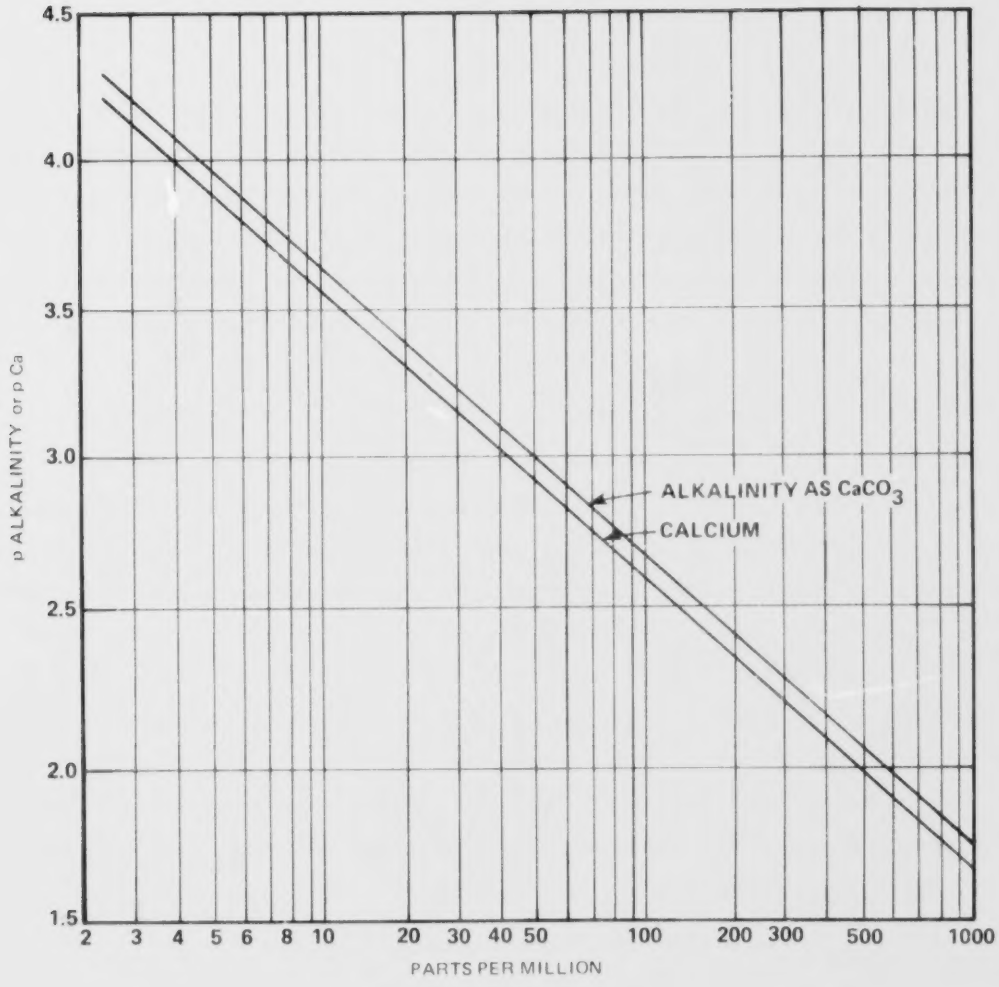

Figure A-1. Conversion from alkalinity or Ca++ to the respective p values.

Table A-1 is entered at the ionic strength of the solution, defined by:

$$\mu = \tfrac{1}{2} \Sigma C_n v_n^2, \text{ where} \qquad (7)$$

μ = ionic strength

C_n = the concentration of ion "n" in moles/liter

v_n = the valence of ion "n"

The Langelier criterion described above is useful for predicting both the scaling and the corrosive tendency of various waters. When L.I. equals zero, the water is stable and there is no tendency for calcium carbonate scale to form. This is equally true for negative values which indicate, in addition, a tendency toward corrosion of metals, particularly the ferrous metals. A positive L.I. signifies decreased corrosiveness and a tendency to deposit calcium carbonate scale.

CALCIUM SULFATE SCALING

The solubility of calcium sulfate is quite limited. In addition, as in the case of calcium carbonate, it has a negative temperature coefficient of solubility. That is, the solubility of calcium sulfate decreases as the temperature of the water rises. Thus it poses the danger of calcium sulfate scale formation in the tubes of the power plant condenser.

The calcium sulfate solubility is one of the constraints on the permissible concentration of the cooling water in the cooling tower loop. In addition to the sulfate ion originally in the feed water to the plant, reaction (4) contributes additional sulfate. One alternative is to soften the feed water by addition of slaked lime:

$$Ca^{++} + 2HCO_3^- + Ca(OH)_2 \rightarrow 2CaCO_3 + 2H_2O \qquad (8)$$

If the water contains calcium ions in excess of the HCO_3^- present, that is, if it contains noncarbonate hardness, soda ash must be added:

$$Ca^{++} + Na_2CO_3 \rightarrow CaCO_3 + 2Na^+ \qquad (9)$$

TABLE A-1. VALUES OF pK_2' AND pK_s' AT 25°C FOR VARIOUS STRENGTHS AND OF THE DIFFERENCE ($pK_2' - pK_s'$) FOR VARIOUS TEMPERATURES (Ref. 6)

IONIC STRENGTH	TOTAL DISSOLVED SOLIDS	25°C.			($pK_2' - pK_s'$)							
		pK_2'	pK_s'	$pK_2' - pK_s'$	0°C.	10°C.	20°C.	50°C.	60°C.	70°C.	80°C.	90°C.
.0000	0	10.26	8.32	1.94	2.20	2.09	1.99	1.73	1.65	1.58	1.51	1.44
.0005	20	10.26	8.23	2.03	2.29	2.18	2.08	1.82	1.74	1.67	1.60	1.53
.001	40	10.26	8.19	2.07	2.33	2.22	2.12	1.86	1.78	1.71	1.64	1.57
.002	80	10.25	8.14	2.11	2.37	2.26	2.16	1.90	1.82	1.75	1.68	1.61
.003	120	10.25	8.10	2.15	2.41	2.30	2.20	1.94	1.86	1.79	1.72	1.65
.004	160	10.24	8.07	2.17	2.43	2.32	2.22	1.96	1.88	1.81	1.74	1.67
.005	200	10.24	8.04	2.20	2.46	2.35	2.25	1.99	1.91	1.84	1.77	1.70
.006	240	10.24	8.01	2.23	2.49	2.38	2.28	2.03	1.94	1.87	1.80	1.73
.007	280	10.23	7.98	2.25	2.51	2.40	2.30	2.05	1.96	1.89	1.82	1.75
.008	320	10.23	7.96	2.27	2.53	2.42	2.32	2.07	1.98	1.91	1.84	1.77
.009	360	10.22	7.94	2.28	2.54	2.43	2.33	2.08	1.99	1.92	1.85	1.78
.010	400	10.22	7.92	2.30	2.56	2.45	2.35	2.10	2.01	1.94	1.87	1.80
.011	440	10.22	7.90	2.32	2.58	2.47	2.37	2.12	2.03	1.96	1.89	1.82
.012	480	10.21	7.88	2.33	2.59	2.49	2.39	2.13	2.04	1.97	1.90	1.83
.013	520	10.21	7.86	2.35	2.61	2.50	2.40	2.15	2.06	1.99	1.92	1.85
.014	560	10.20	7.85	2.36	2.62	2.51	2.41	2.16	2.07	2.00	1.93	1.86
.015	600	10.20	7.83	2.37	2.63	2.52	2.42	2.17	2.08	2.01	1.94	1.87
.016	640	10.20	7.81	2.39	2.65	2.54	2.44	2.19	2.10	2.03	1.96	1.89
.017	680	10.19	7.80	2.40	2.66	2.55	2.45	2.20	2.11	2.04	1.97	1.90
.018	720	10.19	7.78	2.41	2.67	2.56	2.46	2.21	2.12	2.05	1.98	1.91
.019	760	10.18	7.77	2.41	2.67	2.57	2.47	2.21	2.12	2.05	1.98	1.91
.020	800	10.18	7.76	2.42	2.68	2.58	2.48	2.22	2.13	2.06	1.99	1.92

Since the cooling tower water is concentrated almost to the saturation value of calcium sulfate (unless silica is limiting – see Subsection 4.1.3), a second softening step is required before using any blowdown concentration process that cannot tolerate scale.

SILICA SCALING

Although the solubility of silica varies somewhat with the water chemistry, many water technologists arbitrarily set an upper limit of 150 to 200 ppm on silica content. If not already limited by the danger of calcium sulfate scale, the water in the cooling tower basin is generally discharged to waste when its silica content reaches the predetermined value.

To permit evaporation in the cooling tower to a high degree of concentration (frequently referred to as "cycles of concentration"), the feed water is subjected to a silica reduction treatment consisting of coprecipitation of the silica with magnesium hydroxide. The removal of silica by magnesium hydroxide precipitation has been described as a chemical reaction by some investigators, absorption by others. The magnesium can be added in the form of very fine or "activated" magnesium oxide, or the hydroxide can be formed in situ by the precipitation of magnesium ions present in the water, by adding excess lime:

$$Mg^{++} + Ca(OH)_2 \rightarrow Mg(OH)_2 + Ca^{++} \qquad (10)$$

The calcium ions introduced in this process are then removed by reactions (8) and (9).

The softened water is finally acidified to adjust its pH to prevent post-precipitation of calcium carbonate, as discussed in Subsection 4.1.3.

Appendix B

CONVERTING UNITS OF MEASURE

Environmental Protection Agency policy is to express all measurements used in Agency documents in metric units. In this report, however, to avoid undue costs or lack of clarity, English units are used throughout. Conversion factors from English to metric units are given below:

To Convert From	To	Multiply By
$Btu/ft^2 \cdot hr \cdot 1^oF$	$J/m^2 \cdot sec \cdot 1^oK$	5.677
scfm (60^oF)	Nm^3/hr (0^oC)	1.61
cfm	m^3/hr	1.70
^oF	^oC	$(^oF-32)/1.8$
ft	m	0.305
ft/hr	m/hr	0.305
ft/sec	m/sec	0.305
ft^2	m^2	0.0929
ft^2/tons per day	m^2/metric tons per day	0.102
gal/mcf	l/m^3	0.134
gpm	l/min	3.79
gpm/ft^2	$l/min/m^2$	40.8
gr/scf	gm/m^3	2.29
in.	cm	2.54
in. H_2O	mm Hg	1.87
lb	gm	454
lb-moles	gm-moles	454
lb-moles/hr	gm-moles/min	7.56
lb-moles/hr ft^2	gm-moles/min/m^2	81.4
lb-moles/min	gm-moles/sec	7.56
psia	kilopascal	6.895